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ACID HYDROLYSIS OF CELLULOSE IN REFUSE TO SUGAR AND ITS FERMENTATION TO ALCOHOL

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This study reports on the technical and economic feasibility of a process for hydrolyzing the cellulosic content of refuse to sugar. The sugar, in turn, is fermented to produce ethanol. The processes reported on are: (1) the bench-scale kinetics of acid hydrolysis of paper in a batch reactor and a continuous flow isothermal plug reactor; (2) the economic analysis of the acid hydrolysis and; (3) the fermentation of sugar obtained from the hydrolysis of refuse. Under proper circumstances, converting the cellulose in refuse to sugar or ethanol is believed economically feasible.

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FOREWORD

Man and his environment must be protected from the adverse effects of pesticides, radiation, noise and other forms of pollution, and the unwise management of solid waste. Efforts to protect the environment require a focus that recognizes the interplay between the components of our physical environment--air, water, and land. The National Environmental Research Centers provide this multidisciplinary focus through programs engaged in

- studies on the effects of environmental contaminants on man and the biosphere, and
- a search for ways to prevent contamination and to recycle valuable resources.

In an attempt to solve the problem of solid waste disposal, this study examines the hydrolysis of the cellulosic content of refuse to sugar. This grant, published by the National Environmental Research Center, Cincinnati, provides an informative study of the technical and economic feasibility of the acid hydrolysis process.

A.W. Breidenbach, Ph.D.
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National Environmental
Research Center, Cincinnati

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ABSTRACT

The kinetics for the acid hydrolysis of cellulose in paper are experimentally determined over a range of 180°C to 240°C and 0.2% to 1.0% acid in a batch reactor. Since the data fit a sequential first order, irreversible, kinetic model, there is an optimum reaction time where the concentration of sugar for each temperature and acid concentration is maximum. The maximum sugar concentration can be increased by increasing the temperature and/or acid concentration, with a corresponding lower reaction time.

The batch kinetics were verified in a bench scale flow reactor which operated essentially isothermally. Yields of sugar as high as 52 to 54% were obtained. While the feasibility of flow reactors is demonstrated, a large scale pilot plant would be needed to obtain the best solid to liquid ratio and to design the proper pump and pressure regulator for handling a paper slurry.

A flow sheet of a hydrolysis plant is prepared with an economic analysis. In order to show how the design parameters affect the economics, a sensitivity analysis is presented. The major parameters are percent cellulose in the raw refuse and the solid to liquid ratio used in the hydrolysis. The sugar solution is concentrated via a multieffect evaporator to 12%, to be used in a commercial fermentation. As the solid to liquid ratio in hydrolysis is increased, the concentration of sugar is increased, which reduces the evaporator size and cost. Also the higher solid to liquid ratio reduces the heating costs, since more product per unit mass of solution is produced.

The yield of ethanol via fermentation of the sugar from hydrolyzed cellulose with its initial organic impurities has been demonstrated to be insensitive to pH, temperature, sugar concentration, and source of refuse, over the ranges studied. The rate of fermentation is increased with increased sugar concentration and yeast concentration. Commercial fermentation of molasses gives about 80% yield, while yields between 85 to 94% were realized in this study.

The final conclusion is that, under proper circumstances, it can be economically attractive to convert the cellulose in refuse to sugar or ethanol.

1. INTRODUCTION

The current rate of solid waste generation in the United States is about 5 pounds per person per year or roughly 185 million tons per year. The existing methods of solid waste disposal have been criticized for various reasons. Open dump burning is outlawed in many states by the new legislation to prevent air pollution. Sanitary landfill sites cannot be used as building areas because of land settlement and the problem of methane gas generation. Moreover, for large metropolitan areas, local landfill sites do not exist. Incineration involves a comparatively high initial capital investment and the disposal cost to the community is about $4-7 per ton of refuse. New air pollution laws will require future incinerators to be equipped with costly particle removing techniques such as scrubbers, cyclones and electrostatic precipitators. This will further raise the waste disposal costs using incinerators. Composting is practiced in some continental countries but it is not an attractive alternative in the United States as the product has no ready market.

A. Porteous[1] proposed a new technique for waste disposal which would utilize refuse as a raw material. His process consisted of hydrolyzing the cellulosic content of the refuse to sugar. This sugar in turn could be used to produce ethanol by fermentation. He compared the costs of the existing methods of solid waste disposal with the proposed technique mentioned above. Summary of his cost analysis appears in Table 1-1. Because of the encouraging results of this preliminary analysis, a study was undertaken with the support of the Bureau of Solid Waste Management to determine the technical and economic feasibility of such a process.

Kinetics of hydrolysis of paper are presented in Chapter 2. These experiments were done in a non-isothermal batch reactor. A sequential pseudo-first order model, Cellulose $\longrightarrow$ Sugar $\longrightarrow$ Decomposed Sugar, was found to fit the data. Chapter 3 deals with the study of hydrolysis of refuse in a continuous isothermal plug flow reactor. Economic analysis of the acid hydrolysis of refuse, based on the experimental data obtained in Chapters 2 and 3, is presented in Chapter 4. Fermentation of

sugar obtained from the hydrolysis of refuse, together with the treatment of the process effluent, are studied in Chapter 5. Conclusions and recommendations for future work are detailed in Chapter 6.

Table 1-1

COMPARISON OF VARIOUS REFUSE DISPOSAL PROCESSES

PROCESS	PRODUCT	PROFIT $/YEAR	CAPITAL COST PER TON OF DAILY CAPACITY $	REMARKS
Incineration	Disposal	-328,000	8,000	Used as a performance norm.
Destructive Distillation	+ Acetic Acid	-330,000	25,600	Capital cost too high. Marketable product.
Composting	+ Compost	+246,000	10,800	Private venture. Rapid digestion process. Uncertain market.
Hydrolysis	+ Ethanol	+1,142,000	8,520	Profit based on current selling price of 52¢/gal. Ready market.

2. KINETICS OF ACID HYDROLYSIS OF CELLULOSE FOUND IN PAPER REFUSE

2.1 Introduction

Saeman[2] developed a kinetic model for the hydrolysis of cellulose in wood chips. Porteous[1], in his economic analysis, assumed that the model developed by Saeman would be applicable to the cellulosic material in refuse. Porteous based his economic analysis on the extrapolated yield of sugar at 230°C with 0.4 weight % acid concentration. This chapter presents the kinetics of acid hydrolysis of cellulose from paper, based on the experimental data obtained from batch hydrolysis of paper. For details beyond those presented here, the reader is referred to the work of R.D. Fagan[3].

From an economic point of view, it is necessary to carry out the hydrolysis of cellulose in dilute acid at high temperature. Although the reaction is heterogeneous, it can be treated as a homogeneous reaction when the reactant cellulose is dispersed as fine particles of 20 mesh or less. Under these conditions, Saeman[2] showed that the hydrolysis of the crystalline wood cellulose is described by an $A \xrightarrow{K_1} B \xrightarrow{K_2} C$ reaction governed by the following equations.

$$\frac{dC_A}{dt} = -K_1 C_A \qquad (1)$$

$$\frac{dC_B}{dt} = -K_2 C_B + K_1 C_A \qquad (2)$$

where C_A = concentration of cellulose

C_B = concentration of sugar

K_1 = reaction rate constant for cellulose to sugar reaction

K_2 = reaction rate constant for sugar to decomposition products

t = time

$$K_i = P_i e^{-\Delta H_i / RT}, \quad i = 1, 2 \qquad (3)$$

Since Saeman was able to model the hydrolysis of cellulose in wood chips very effectively with the above equations. it seemed reasonable that cellulose in paper would have different rate constants but effectively follow the same overall kinetic scheme. In fact, Saeman showed that cellulose from cotton linters followed the same kinetic scheme but had different rate constants than the cellulose from wood ships.

2.2 Method

Normally, kinetic data are obtained under isothermal conditions, but above 200°C the maximum sugar concentration occurs in a fraction of a minute. As a result, the heat-up time cannot be neglected with respect to the reaction time. While a procedure for isothermal measurements up to 250°C has been developed[4], the non-isothermal procedure used here assumes that the kinetic model is correct and only the parameters need to be estimated.

Sugar concentrations were measured by an O-toluidine, colorimetric method[3]. The test, while sensitive to the presence of metal (Fe) ions, was not affected by the presence of decomposed sugars. Cellulose or potential sugar content was determined by a sugar determination following a quantitative (high acid concentration, low temperature) saccharification[2].

An 0.5 gram sample of paper or refuse,which had been run through a Wiley ball mill to produce 2 mm particles, was placed in a glass culture tube with 20 ml of acid solution. This vessel was placed in a 1" x 4" stainless steel nipple and sealed to act as an autoclave. The cap of the clave was taped to accommodate a thermocouple well connector. The clave was placed in a constant temperature oil bath which had been set at 260°C and the temperature-time history was recorded. The claves were removed from the bath at various intervals, quenched in ice water, and analyzed for sugar content. Six to eleven time samples were taken for each acid concentration (0.2%, 0.5%, and 1.0% by weight sulfuric acid).

The initial conditions for each run were obtained from a clave quenched when its temperature reached 180°C and analyzed for both sugar and residual potential sugar (cellulose) content. The data obtained from such a procedure consisted of 1) quantitative potential sugar (cellulose content) at 180°C, 2) sugar yield at intervals of time, and 3) temperature-time history.

With measured initial conditions on C_A and C_B at 180°C, a temperature-time function derived from a 4th order polynomial curve fit to the experimental temperature profile, and assigned values for P_1, ΔH_1, P_2, ΔH_2, equations (1) through (3) were numerically integrated for the sugar yield as a function of time. It was necessary to search for the values of P_1, ΔH_1, P_2, and ΔH_2, which resulted in yields of sugar which best agreed with those obtained experimentally. A Hooke-Jeeves pattern search[5] and a Runge-Kutta numerical analysis were used to determine the rate constant parameters which minimized the sum of the square errors between the predicted and the experimental sugar yields.

2.3 Results

The potential sugar of cellulose content of the materials used for the acid hydrolysis is shown in Table 2-1. The Kraft paper results compare well with those reported in the literature[4]. The yields (weight of sugar obtained per weight of cellulose changed) as a function of temperature and time histories are shown in Figure 2-1 for samples of milled Kraft paper. The actual experimental data points are compared with values computed from the model in equations 1, 2, and 3.

A computer program was used to perform the search which determined the pre-exponential factors and activation energies that would best fit the data for the 0.5% acid runs. It was then assumed that these activation energies could be used for the 0.2% and 1% acid experiments, thus only requiring a search on the two pre-exponential parameters for this data. The estimated parameter values are presented in Table 2-2.

Table 2-1

POTENTIAL SUGAR CONTENT

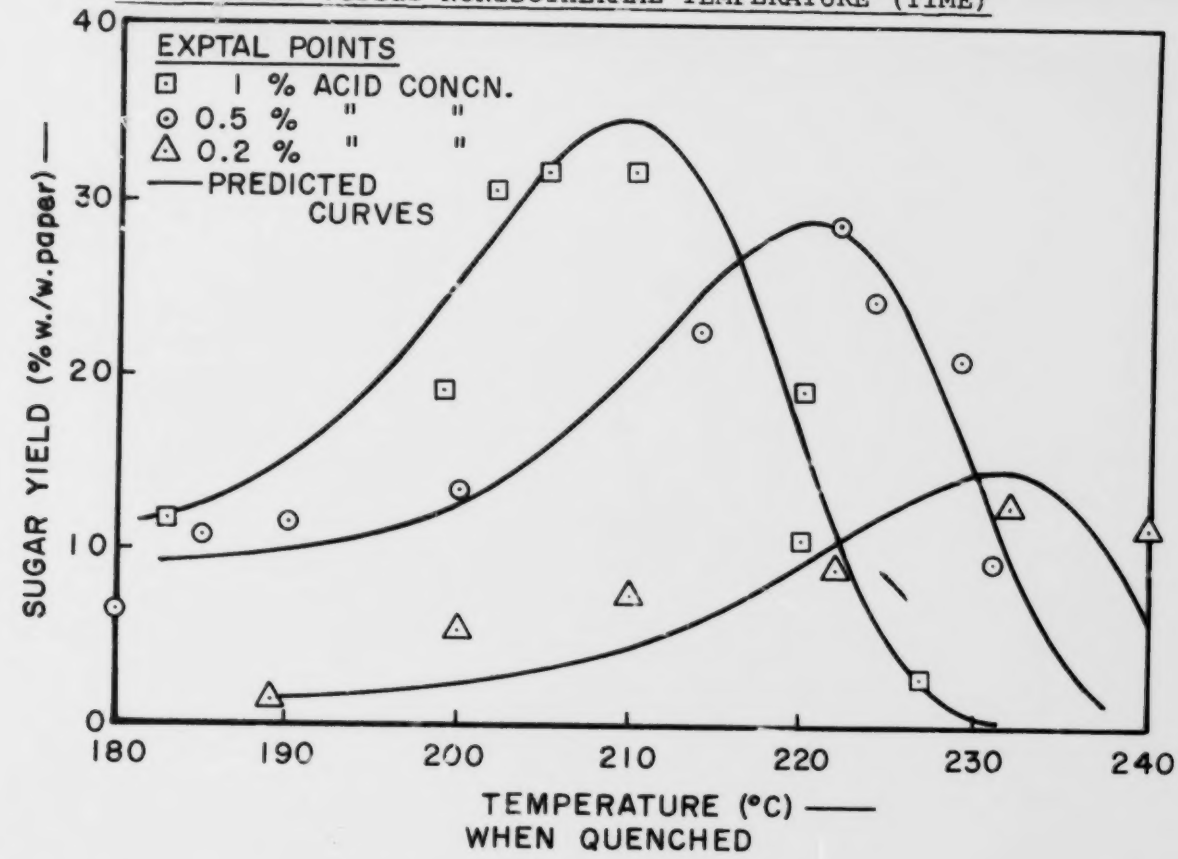
Sample	Weight Sugar/Weight Paper %
Ground Kraft Paper	86, 84
Slurried Kraft Paper	84.8, 80.5
Ground Refuse #1	38.0, 38.4
Ground Refuse #2	52.6, 52.6

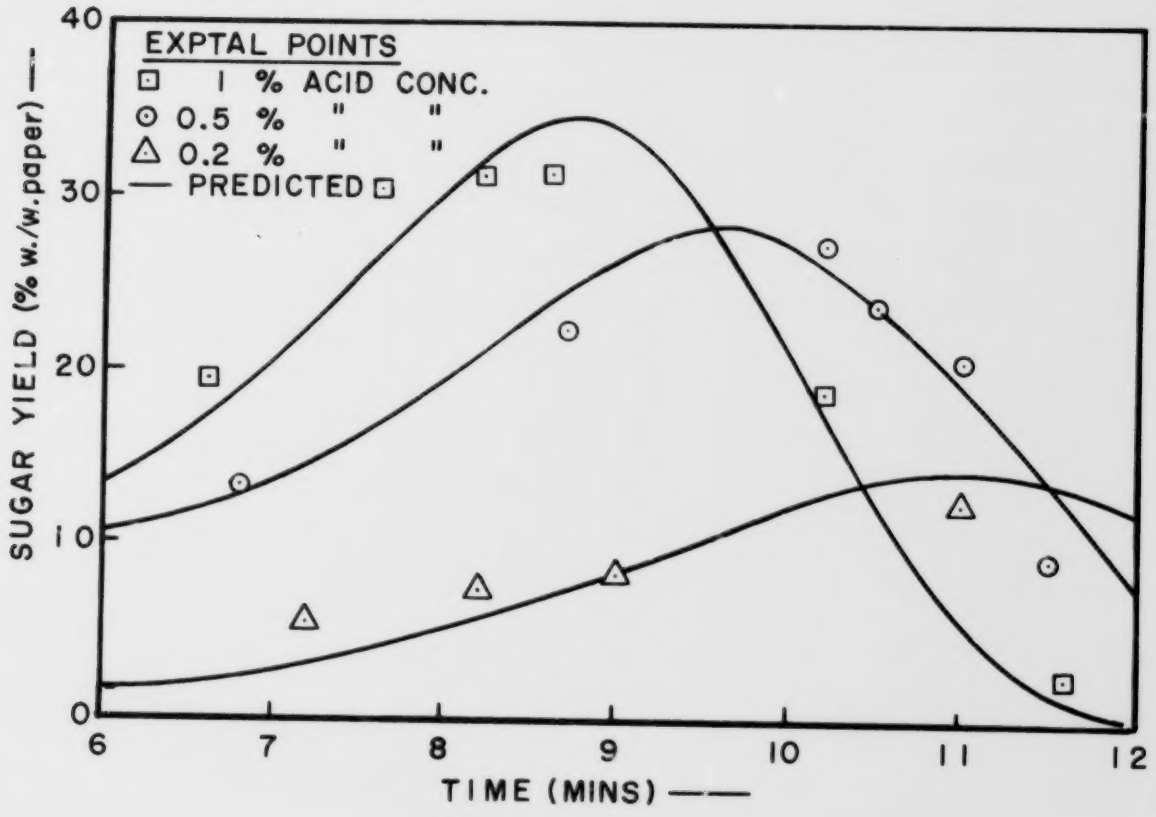
Table 2-2

KINETIC PARAMETER FOR PAPER HYDROLYSIS

	Rate Constant for Hydrolysis K_1		Rate Constant for Sugar Decomposition K_2	
% Acid	Pre-Exponential Factor P_1 min^{-1}	Activation Energy ∇H_1 K-cal/mole	Pre-Exponential Factor P_2 min^{-1}	Activation Energy ∇H_2 K-cal/mole
0.2%	1.661×10^{19}		2.21×10^{14}	
0.5%	7.59×10^{19}	45,100	2.67×10^{14}	32,800
1.0%	2.95×10^{20}		5.57×10^{14}	

Figure 2-1
SUGAR YIELD versus NONISOTHERMAL TEMPERATURE (TIME)





In an acid catalyzed reaction, the overall reaction rate is usually proportional to the acid concentration as well as the reactant concentration. Although this would normally result in second order kinetics, the fact that the acid concentration is constant allows the acid concentration to be expressed as part of the pre-exponential factor of Arrhenius equations, so that $P_1 = P_0 C^m$ where P_1 is the pre-exponential of K_1 in $K_1 = P_1 e^{-\Delta H_1/RT}$. A similar expression applies to K_2. The parameters M and P_0 are obtained from the log of the pre-exponential factors versus the log of the acid concentration. The estimated parameters are given in equations 4 and 5.

When the mean square for experimental error (obtained by replicate runs) is compared to the mean square for deviations between the model and observation in an F test, the lack of fit of the model is of the same order as the experimental error. Thus the model fits the data.

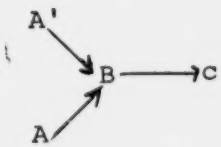
A question concerning the accuracy of the experimental data arises from the ice water quench time. It was found that approximately 20 seconds are required to lower the reactor temperature from 230°C to 170°C. Using the reaction rates calculated for a 1% acid solution, a numerical integration was performed for a linear drop in temperature from 230 to 170°C with an initial sugar concentration of 11%. This calculation revealed that approximately a 1% loss in sugar yield could be expected due to the quench. Since the condition used for this test is the most severe encountered, the error due to quench time was ignored.

Since the proposed hydrolysis plant will operate with a paper slurry, it was necessary to determine whether mass transfer effects would change the rate constants if fibrous material were used in place of ground particles. A sample of Kraft paper was pulped with water in a standard Waring blender. This sample was then dried and 0.5 gram samples were prepared from the fibrous sheet. These samples were mixed with 20 ml. of 0.5% acid producing a fibrous slurry which was then hydrolyzed. The

agreement between predicted and experimental results (see Figure 2-2) indicates that the kinetic relationship holds for fibrous samples as well as ground samples. It is therefore possible to predict isothermal yields for plant operation using a pulped paper slurry.

Referring again to Figure 2-1, it is noted that approximately 6 to 11% sugar yields were obtained at 180°C. From the rate constant for hydrolysis, yields of this magnitude, at low temperatures and short heat-up, should not be obtained. The discrepancy can be explained, since the same effect occurs with wood chips[2] by the different types of cellulose hydrolyzed.

There is amorphous (A') and crystalline (A) cellulose present, and both decompose to yield sugars. Hence we have



The reaction, $A' \longrightarrow B$, goes much faster than the other reactions, and at 180°C it is essentially instantaneous. Hence, only the $A \longrightarrow B \longrightarrow C$ reactions were observed in this study. Consequently, one cannot take the initial concentration of B to be zero because of the $A' \longrightarrow B$ reaction, and for this reason the initial concentrations were measured. Naturally, the kinetic parameters estimated here apply only for the residual cellulose.

With a satisfactory model for paper hydrolysis, we turn now to our original interest in obtaining the maximum sugar yield. The ratio of K_1/K_2 increases with acid concentration and the temperature. The predicted isothermal yields of sugar from paper cellulose are shown in Figure 2-3 as a function of time. It was assumed that 10% of the original potential sugar yield would hydrolyze instantaneously. These curves give the residence times necessary to obtain maximum yield at various temperatures and acid concentrations.

Figure 2-2

SUGAR YIELD FROM (KRAFT FIBERS)
versus TEMPERATURE

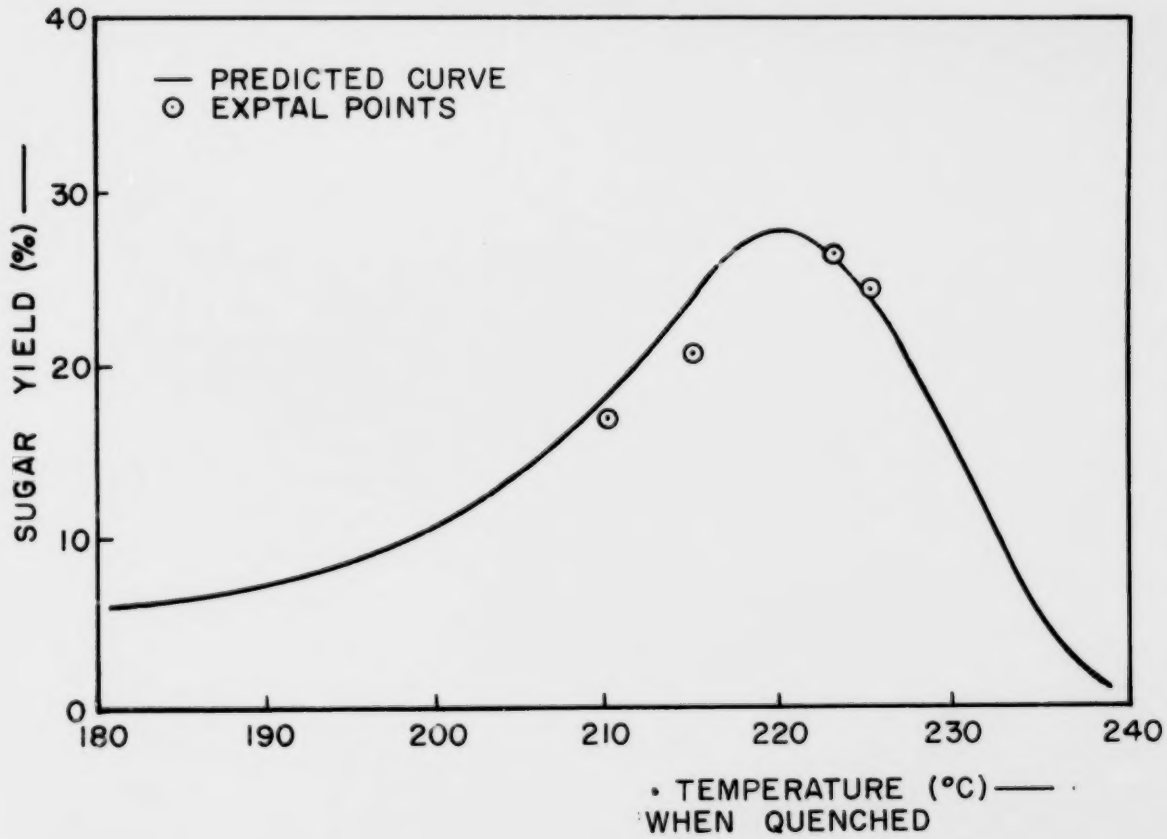
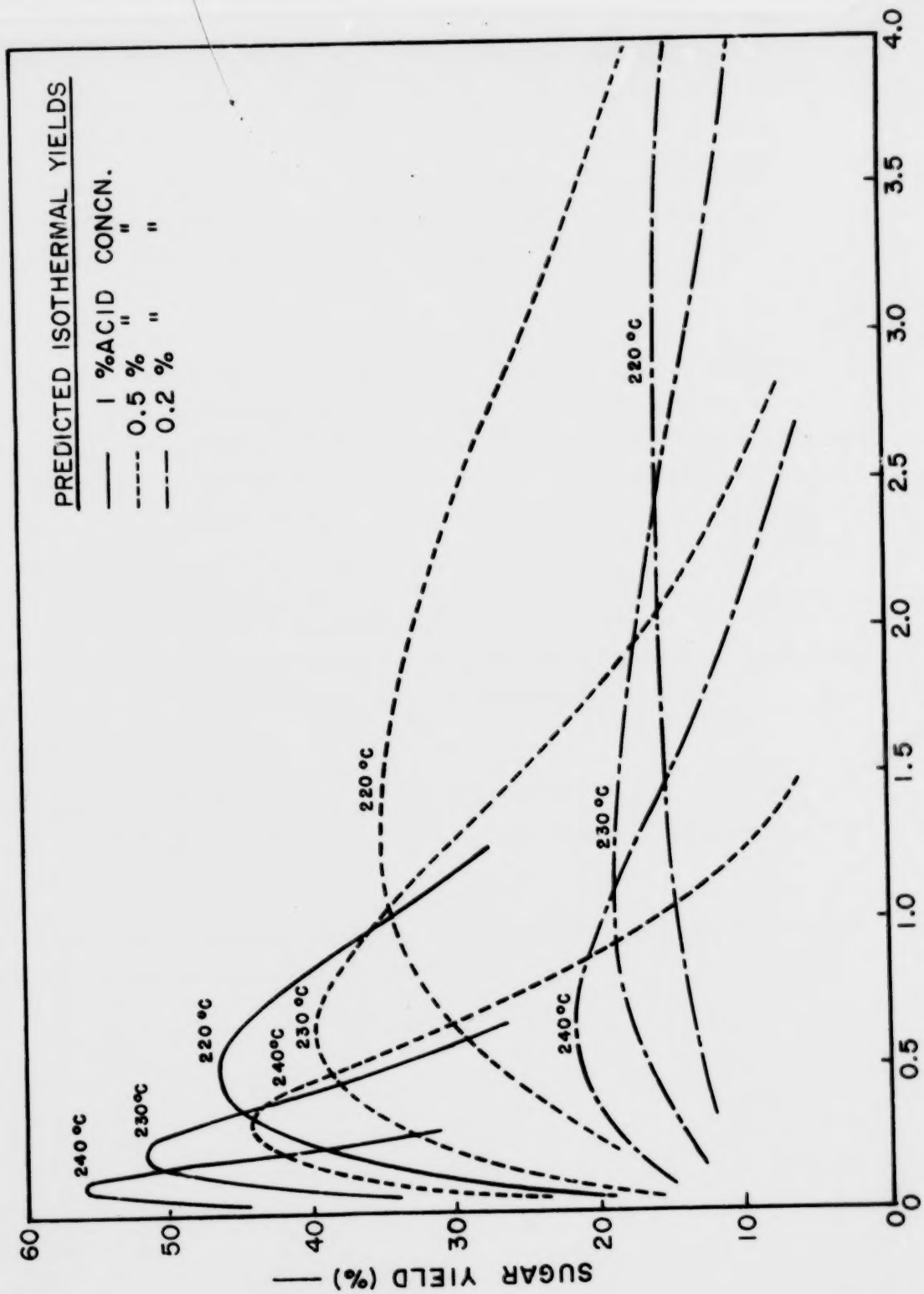


Figure 2-3

PREDICTED ISOTHERMAL SUGARS versus TIME



Since the yield is reported as a percentage of the potential sugar content, the actual weight of sugar at maximum yield is obtained once the potential sugar content of the refuse is known by quantitative saccharification. The practical limit on temperature is about 230°C because the reaction time for optimum yield is about 20 seconds.

The samples of ground refuse in Table 2-1 gave sugar yields as predicted from the model. Thus, the impurities in the refuse do not interfere with the hydrolysis.

2.4 Conclusions

It was shown by the experimental analysis that paper cellulose does hydrolyze to sugar, and that the hydrolysis reaction could be modeled by the $A \longrightarrow B \longrightarrow C$ irreversible reaction model used by Saeman[2]. The reaction rate was found to be represented by:

Decomposition of Cellulose ($A \longrightarrow B$):

$$r_1 = 28 \times 10^{19} C^{1.78} \exp(-45.100/RT) C_A \qquad (4)$$

where C = cellulose concentration, wt. of residual potential sugar/wt. of slurry

Decomposition of Glucose ($B \longrightarrow C$):

$$r_2 = 4.9 \times 10^{14} C^{.555} \exp(-32,800/RT) C_B \qquad (5)$$

where C_B = sugar concentration, wt. of sugar/wt. of slurry

The sugar yield depends on time, temperature, and acid concentration in the manner shown in Figure 2-3.

3. HYDROLYSIS OF REFUSE IN A FLOW REACTOR

3.1 Introduction

Experimental work done on the acid hydrolysis of refuse in this lab showed that a sugar yield of 51 weight % (based on the potential sugar in the refuse) can be obtained if the hydrolysis reaction is carried out isothermally at 230°C using 1 weight % sulfuric acid and having a reaction time of 20 seconds (Figure 2-3). This conclusion is based on a kinetic model developed from the data obtained from non-isothermal batch hydrolysis of refuse. The sugar yield is a function of reaction temperature, time of reaction, and the concentration of sulfuric acid. Based on this kinetic data, it was decided to build a bench scale flow reactor which would hydrolyze refuse slurry. The kinetics showed that, given an acid concentration, the highest sugar yield is obtained by operating at the highest achievable isothermal temperature with its corresponding residence time. By operating at steady state, an isothermal reaction can be carried out for any desired reaction time by controlling the flow rate through the reactor. The objectives of obtaining experimental data on a flow reactor were: i) to check whether the optimum yield predicted from the batch kinetic data was actually attainable in a flow reactor, ii) to compare the kinetics of the flow reactor with the predicted batch kinetic data, and iii) to observe what types of scale-up problems need to be considered, such as mixing the reactants, pumping the slurry, and quenching the reaction at the desired time.

3.2 Equipment

A schematic diagram of the flow reactor is given in Figure 3-1. A water refuse slurry from the storage tank, (1), is pumped to a heater, (6) by a positive displacement pump, (3). Hot slurry coming out of the heater mixes with the concentrated sulfuric acid in a mixing 'tee', (9) just prior to entering the reactor, (10). Pressure in the reactor was controlled by a valve, (18) which was operated manually.

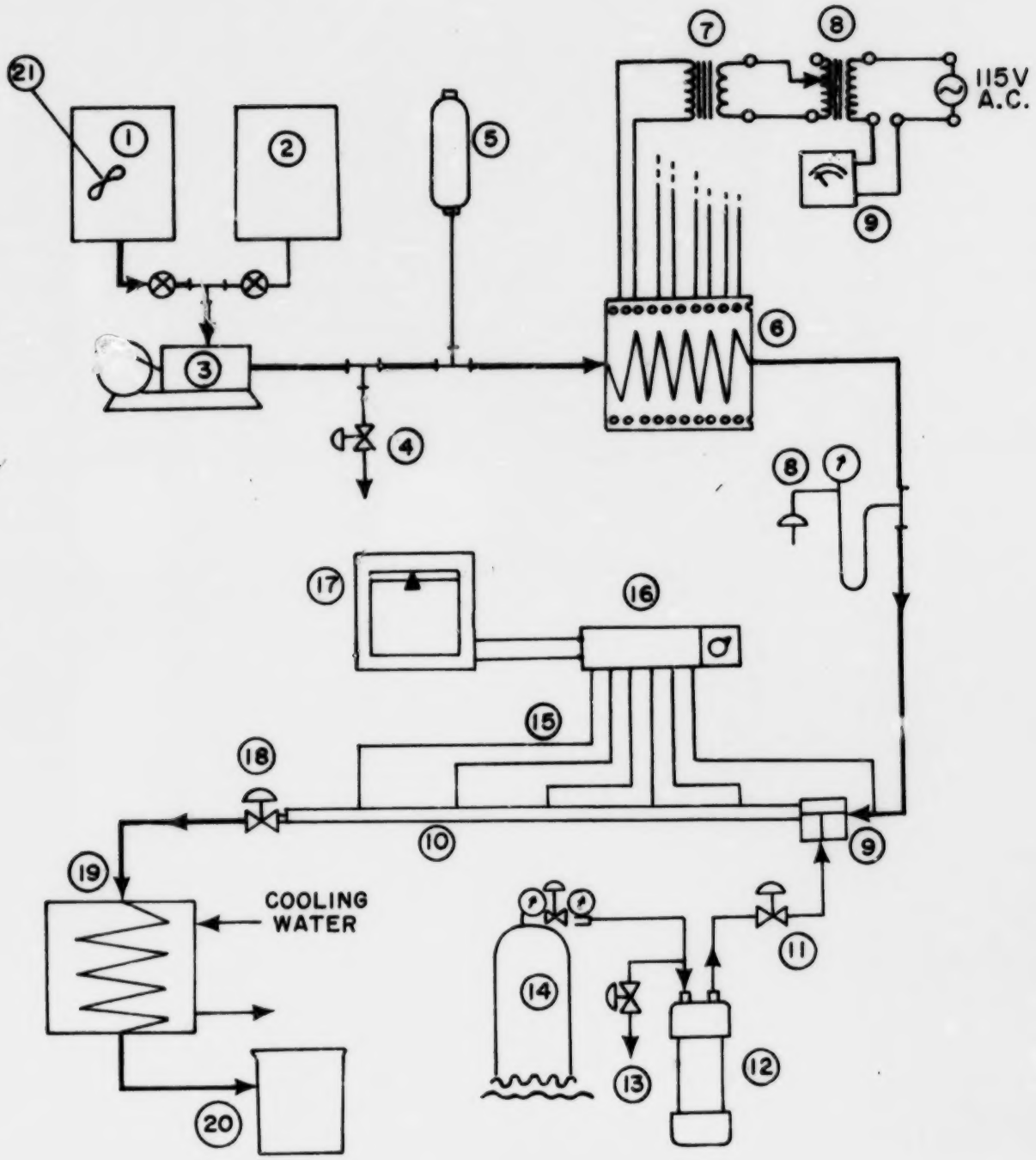


Figure 3-1. FLOW REACTOR -- SCHEMATIC DIAGRAM

Products of hydrolysis were cooled in a quencher, (19) and collected in the product accumulator, (20).

Slurry storage tank was a fifteen liter cylinderical vessel made of PVC. The slurry was kept in a homogeneous suspension with a stirrer, (21) mounted at the top of the tank. A water storage tank, (2) was also connected to the pump inlet. At the beginning of an experiment, water was pumped through the reactor until the operating conditions of temperature and pressure were stabilized at the desired level. At this stage, the slurry storage tank was connected to the pump, disconnecting the water storage tank. Another function of the water storage tank was to flush the entire system (the pump, the heater, the reactor, and the quencher) at the end of a run to avoid settling of the residual slurry causing choking at the valves and the joints.

The pump, (3) was a 'MILROYAL' diaphragm pump. The pumping rate could be varied from 0.1 to 6.0 gal/hour by adjusting the stroke length. The designed outlet pressure of the pump was 800 psig. The pumping rate was calibrated as a function of the stroke length before the experiments were done. The heater was a 30-feet long tubing of 1/4 inch Carpenter-20 steel. The tube was put in the form of three concentric helixes. The cold stream of the liquid entered the heater in the outer helix and the heated liquid left the heater from the inner helix. The heating of the tube was done by wrapping electric tapes around the tube. The designed tube wall temperature for these electric tapes was 750-800°F. Concentrated sulfuric acid (99.6%) was injected into the stream of hot slurry just before it entered the reactor. In this way, the refuse slurry could be raised to the reaction temperature without having the hydrolysis occur until the acid is added. The heating of slurry can cause thermal decomposition of the cellulose, but this is not significant if kept below 15 minutes. The acid was stored in an acid 'egg', (12). The acid egg was connected to a high pressure nitrogen tank, (14). The flow rate of acid entering the slurry stream was controlled manually by adjusting a micrometering valve,(11)

located upstream of the mixing tee. The reactor was a five feet long tube made of 1/4 inch Carpenter-20 steel. Wall temperature of the reactor at five locations along the length of the reactor was indicated by a multipoint temperature indicator recorder, (17). Pressure in the reactor inlet was indicated by a pressure gauge. A safety rupture disc, (8) was installed in between the heater outlet and the reactor inlet. This disc was designed to rupture when the pressure in the system exceeded 600 psi. The entire tubing, from the inlet of the heater to the outlet of the reactor, was insulated to minimize heat losses.

The valve, (18), situated at the end of the reactor section, was a micrometering needle valve designed to take a pressure drop of 500 psi. This valve was operated manually during a run in order to control the pressure in the reactor. The slurry coming out of this valve entered the quenching section, (19). The quencher consisted of a 10-feet helical coil immersed in ice. The hydrolyzate coming out of the quencher at 25 - 30°C was collected in the product accumulator, (20).

3.3 Experimental Procedure

The slurry was prepared in the slurry storage tank by mixing 100 g of ground refuse with 9.9 liters of water. This gave a solid concentration of 1 weight % in the slurry. Water was pumped through the reactor and the stroke length of the pump adjusted so as to give a residence time in the reactor to a desired level -- usually between 20 to 100 seconds. The reactor outlet valve was then adjusted to build a pressure of 480 psig in the system. The electric heat was then switched on and controlled in such a way to give a desired temperature -- usually between 210 and 235°C -- in the reactor. The temperature drop along the reactor varied between 4 and 6°C for a run. It took about 30 minutes to reach the steady operating conditions of temperature and pressure in the reactor. The slurry was then pumped through the

reactor, disconnecting the water tank. Sulfuric acid was simultaneously introduced in the mixing tee. The flow rate of acid was adjusted to give an acid concentration of about 1 weight % in the slurry. It took about 30 minutes to stabilize the flows of sulfuric acid and the slurry because of the manual adjustments of the valves.

The temperature at various points along the reactor was indicated by a multipoint temperature indicator-recorder. The acid concentration in the hydrolyzate was determined by titrating the sample against a standard NaOH solution. After the flow rates of the slurry and acid had been adjusted as desired and stabilized, hydrolyzate samples were drawn at an interval of ten minutes. Volume of every sample was recorded by collecting the samples in a measuring cylinder for 45 to 90 seconds. This gave the mean residence time for that sample. These samples were later analyzed for acid concentration and sugar concentration[3]. At the end of each run, flows of acid and slurry to the reactor were stopped and the electric heat to the heater was cut off. The system was then flushed with water for about 30 minutes to ensure that all the residual solids and acid were removed from the system.

3.4 Results

The yield of sugar obtained from the hydrolysis of refuse is a function of the temperature of reaction, the residence time, and the acid concentration, as stated in Chapter 2. In the experiments on the flow reactor, there was a temperature drop of 4 to 6°C along the length of the reactor. Hence the reaction temperature was taken to be the mean temperature along the length of the reactor. While this reaction was not exactly isothermal, the reaction temperature varies only 4 to 6°C during a run, which is a great deal closer to the ideal isothermal condition than can be achieved in a batch reactor where the temperature of the reactants changes from room temperature to 230°C in 20 minutes. Residence time was obtained from the observed flow rate at the time of the sampling. The acid concentration was determined by titration. Potential

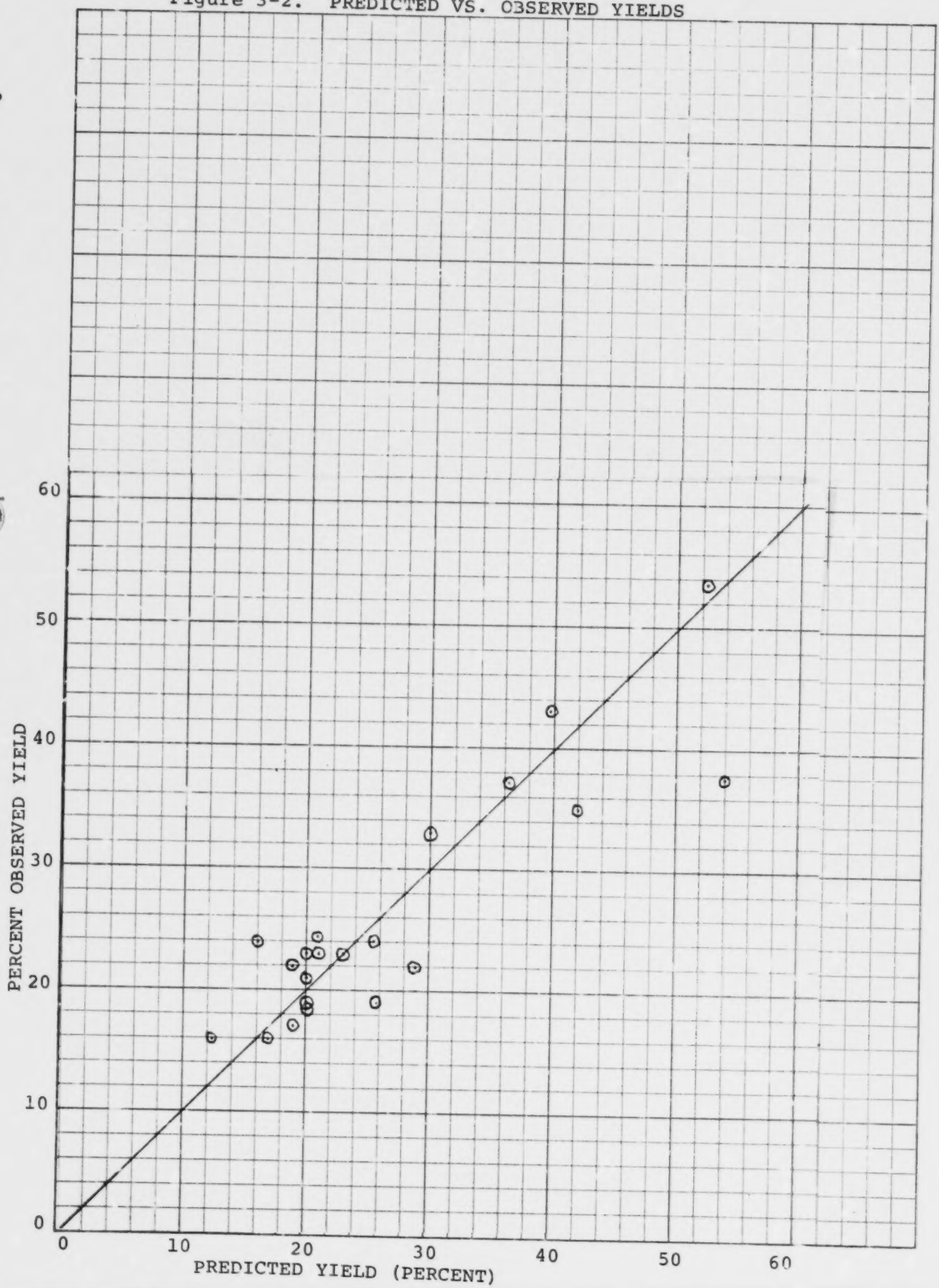
sugar in the sample prior to hydrolysis was calculated from the results of the quantitative saccharification of the ground refuse. Predicted yields for each sample, based on the kinetic model developed in Chapter 2, were calculated from the program REACTAN as a function of temperature, residence time, and the acid concentration. Observed sugar yields were calculated by analyzing the samples for sugar. Table 3-1 gives the observed yields and the predicted yields for the given operating conditions. Figure 3-2 plots the observed yields against the predicted yields.

In calculating the predicted yield, average temperature along the length of the reactor was used. Since the dependence of yield on temperature is far from linear, averaging of the temperature involves a certain amount of error. Moreover, a small variation in the temperature appreciably affects the sugar yield when the operating temperature is high (220 - 235°C). Flow of acid and slurry is difficult to set at prior levels and keep them steady because of the low flow rates involved, the high pressure of operation, and the momentary clogging of solids in the needle valve. For example, pressure fluctuations caused by the quench valve directly affect the temperature. The temperature will always drop to a new saturation temperature if the initial pressure drop lowers the pressure below the saturation pressure of the corresponding initial temperature.

Many of the operating problems will be alleviated if the scale of operation is increased. The reactor temperature will be more isothermal in a large unit where the surface to volume ratio of the reactor will be lower, resulting in a lower heat loss per unit volume. The flow of a two phase slurry through larger valves will not clog as readily as in the needle valve. In fact, a special back pressure regulator can be used with large flow rates. Pumping a slurry in a positive displacement pump which larger inlet and outlet check valves will also be less sensitive to slurry clogging when operating at 1% slurry concentration.

These results do show that a short residence time flow reactor operating essentially isothermal gives sugar yields consistent with the kinetic model developed in Chapter 2. The

Figure 3-2. PREDICTED VS. OBSERVED YIELDS



NO. 312. 5 DIVISIONS PER INCH BOTH WAYS. 25 BY 50 DIVISIONS.
CODEX BOOK COMPANY, INC. NORWOOD, MASSACHUSETTS.
PRINTED IN U.S.A.

Table 3-1

OPERATING CONDITIONS, OBSERVED YIELDS, AND PREDICTED YIELDS

RUN

REACTAN 25 MAY 71 11:39

TEMP.DEG.C	%ACID CONC.	RES.TIME,MIN	OBS.YIELD	PRE.YIELD
211	0.52	0.35	22	19.0623
214	0.45	0.935	19.1	20.2332
214	1.52	0.667	37.5	54.901
215	0.47	0.85	23	22.9993
216.5	0.495	0.513	19.9	18.5264
217	0.39	0.935	23	20.1297
217	0.42	0.875	23	21.404
221	0.531	0.611	33.2	29.7715
222	0.31	0.945	17.	13.8491
222	0.4	0.825	24	25.6502
222	0.4	0.835	19	25.7792
224	0.3	0.821	21	19.305
225	1.07	0.269	53.5	52.605
226.5	0.275	0.553	23.8	16.3961
227	0.335	0.5	24.4	20.8946
227	0.52	0.845	37	36.4153
228	0.36	0.87	22	29.1627
229	0.21	0.525	16	12.6069
229	0.8	0.533	35	42.0696
231.5	1.025	0.467	43	39.5209
232	0.19	0.35	16	17.0011

0.518 SEC. 15 I/O
READY

Table 3-1 (continued)

```
READY

LIS

REACTAN   26 MAY 71   11:37

100 'THIS PRØGRAM ANALYSISES THE DATA FRØM THE FLØW REACTØR
110 ' EXPERIMENTS ØN THE CØNVERSIØN ØF THE SØLID WASTE TØ SUGAR.
120 '
121 READ A1,N1,E1,A2,N2,E2
125 DATA 23.21,1.78,45127,4.905,0.555,32800
130 PRINT" TEMP.DEG.C","%ACID CØNC.","RES.TIME,MIN","ØBS.YIELD","PRE.YI
132 PRINT
140 DIM X(50),Y(50)
150 'DATA INPUT : TEMP.,%ACID,%SUGAR YIELD,RES.TIM.(MINS.)
160 DATA 211,.52,22,.850
170 DATA 214,.45,19.1,.935
180 DATA 214,1.52,37.5,.667
190 DATA 216,.47,23,.850
200 DATA 216.5,.495,19.9,.513
210 DATA 217,.39,23,.935
220 DATA 217,.42,23,.875
230 DATA 221,.531,33.2,.611
240 DATA 222,.31,17,.845
250 DATA 222,.40,24,.825
260 DATA 222,.4,19,.835
270 DATA 224,.30,21,.821
280 DATA 226,1.07,53.5,.269
290 DATA 226.5,.275,23.3,.563
300 DATA 227,.335,24.4,.5
310 DATA 227,.52,37,.845
320 DATA 228,.36,22,.870
330 DATA 229,.21,16,.525
340 DATA 229,.3,35,.588
350 DATA 231.5,1.025,43,.467
370 DATA 232,.19,16,.850
530 DATA 0,0,0,0
600 READ T1,C9,Y9,T9
610 IF T1=0 THEN 999 'END ØF PRØGRAM
620 LET K1=A1*C9↑N1*EXP(-E1/(1.98*(T1+273)))*1E19
630 LET K2=A2*C9↑N2*EXP(-E2/(1.98*(T1+273)))*1E14
640 LET Y5=(K1/(K2-K1))*(EXP(-K1*T9)-EXP(-K2*T9))
650 PRINT
660 PRINT T1,C9,T9,Y9,Y5*100
700 GØ TØ 600
999 END
READY
```

large variation of the actual yields around the expected yields is due to the non-steady temperature and flow rates as pointed out above. It is important to note that yields as high as 53% are achieved (at 226°C, 1.07% acid and 0.269 minutes) which is close to Fagan's prediction of 51% at 230°C, 1% acid and 0.25 minutes. Since it was difficult to set the flow rate of acid and slurry and the operating temperature a priori, it was not possible to operate precisely at the optimum temperature-residence time combination shown in Figure 2-3.

3.5 Conclusions

Table 3-1 shows that hydrolysis of refuse in a continuous isothermal plug flow reactor, operating at high temperatures and low reaction times, gives yields consistent with the yields predicted by the kinetic model developed in Chapter 2. Operating difficulties experienced in this bench scale flow reactor could be minimized if the scale of operation is increased. The results also show that it is possible to heat up a 1 weight % slurry of refuse in water to 230°C in a reasonably short time to avoid thermal degradation of cellulose. There was no difficulty in mixing concentrated acid with 1 weight % slurry in a continuous manner. Effect of thicker slurry on the heat-up time and mixing of acid will have to be studied.

4. ECONOMIC ANALYSIS

The experimental results obtained from the bench scale flow reactor (as described in Chapter 3) and recent cost figures for an operational pretreatment system have been used to facilitate a more detailed cost analysis of the acid hydrolysis process as studied by Fagan[3]. A mass balance simulation program has been written which estimates the various costs involved for the hydrolysis plant operation. These new cost estimations verify the feasibility of the continuous flow reactor system. The previously obtained kinetic parameters are used to calculate the sugar yields for the continuous hydrolysis flow reactor.

4.1 Plant Design

Fagan[3] has demonstrated the desirability of a single stage continuous flow reactor with recycle capabilities. The basic plant design, therefore, is identical to that proposed in his study. The design calculations for the individual components are identical with those used in the previous study. This allows for equitable cost comparisons between the present process simulation routine and that proposed by Fagan[3].

The hydrolysis plant can be divided into three primary stages of operation: 1) separation and pretreatment, ii) hydrolysis and recycle, and iii) neutralization and concentration. The first stage receives the raw municipal waste, separates the main portion of the unhydrolyzable material, and prepares the cellulosic content for the reaction stage. This preparation involves pulping and slurrying the paper content of the refuse. The hydrolysis stage consists of a tubular flow reactor which hydrolyzes the cellulosic material upon the injection of sulfuric acid. Unreacted cellulose may then be recycled through the reactor for optimal conversion to sugar. The acidic product stream from the reactor is neutralized with a lime slurry and the product sugars are concentrated in a feed forward multi-effect evaporator system.

Inputs for the hydrolysis plant consist of the refuse, process water, sulfuric acid, and lime; the outputs are the product sugar solution, the separables from the pretreatment stage, and the hydrolysis waste.

4.1.1 Feed and Pretreatment

Specific cost estimates have been obtained on two refuse pretreatment processes which have been developed by the Black Clawson Company[6]. It has been assumed that a pretreatment process of these general specifications will be used in conjunction with the acid hydrolysis plant.

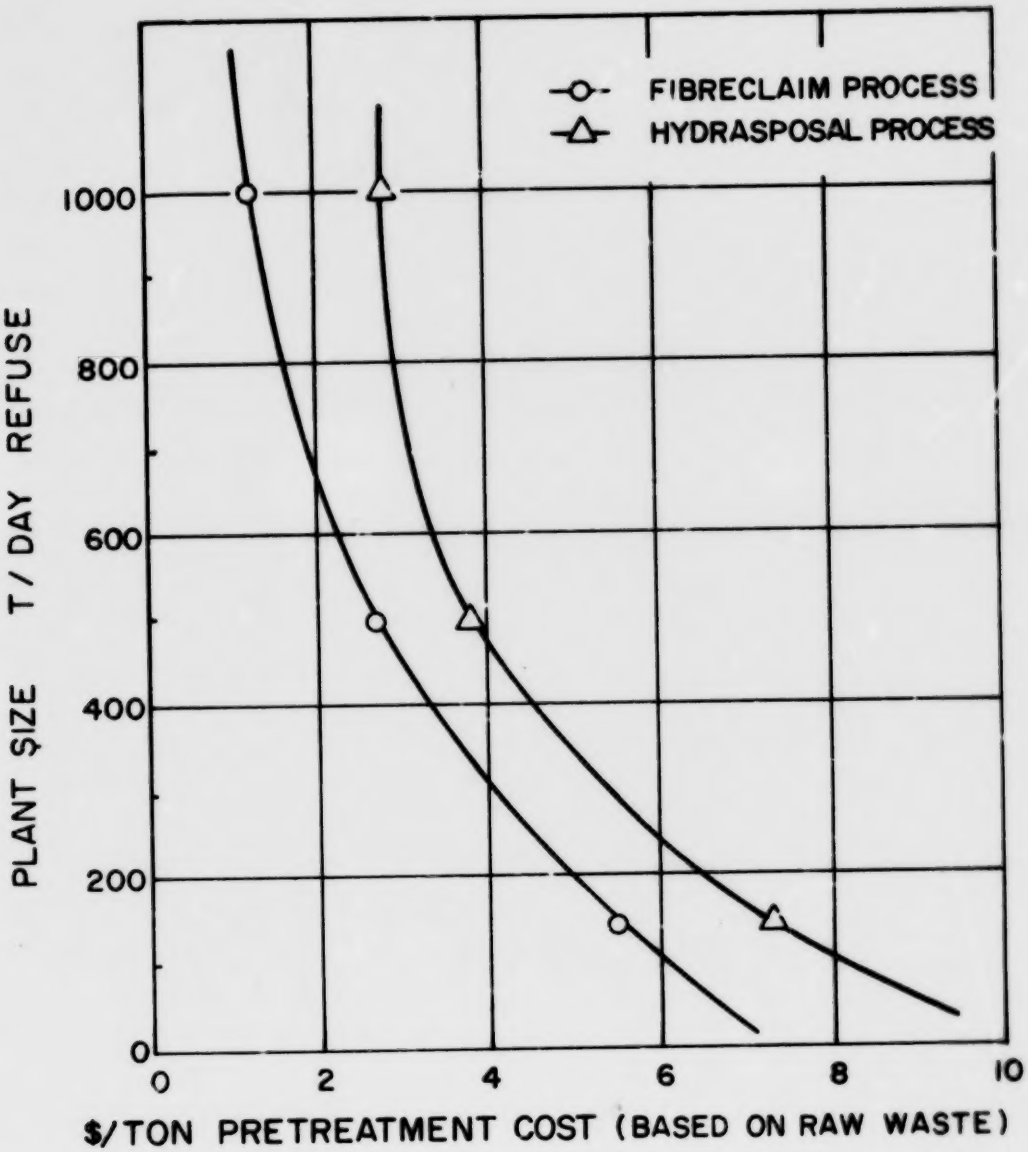
The Black Clawson "FIBRECLAIM" process isolates and removes the long cellulosic fiber which is reusable for the paper manufacture. Metals, glass and other inert organics are removed and the remaining material is suitable for acid hydrolysis. The "Hydrasposal" system is used in conjunction with the "FIBRECLAIM" process as the initial separation stage. This segment of the operation removes all junk metal and the inert inorganic material such as glass and sand. The remainder of the refuse is fed to the hydrapulper which disintegrates all the pulpable and grindable material creating a water slurry with a moisture content of about 50%. The "Hydrasposal" system could be used by itself as the pretreatment stage for the acid hydrolysis plant, thereby the entire cellulosic content of the refuse would be available as feed for the hydrolysis reactor. The "FIBRECLAIM" and "Hydrasposal" systems have demonstrated under pilot plant conditions at the Black Clawson facilities in Middleton, Ohio.

The Black Clawson Company has supplied the cost data which were used to develop the graph in Figure 4-1. The costs shown for a given plant capacity include the entire manufacturing cost per ton of pretreated refuse. Although it is more costly, the "Hydrasposal" process has proven to be more desirable as the pretreatment stage for the hydrolysis process. This benefit is explained by the fact that the entire cellulosic content in the raw waste is retained for the reactor feedstock, whereas in the "FIBRECLAIM" process much of the cellulose is removed to be reused by the paper industry. Therefore, the "Hydrasposal" pretreatment allows greater product sugar yields per ton of initial raw refuse.

Figure 4-2 is a schematic flow diagram of a "Hydrasposal"-type process. Figure 4-3 shows the average composition of raw waste entering the pretreatment stage and the output feedstock for the reactor stage.

Figure 4-1

RETREATMENT PLANT CAPACITY vs. COST PER TON



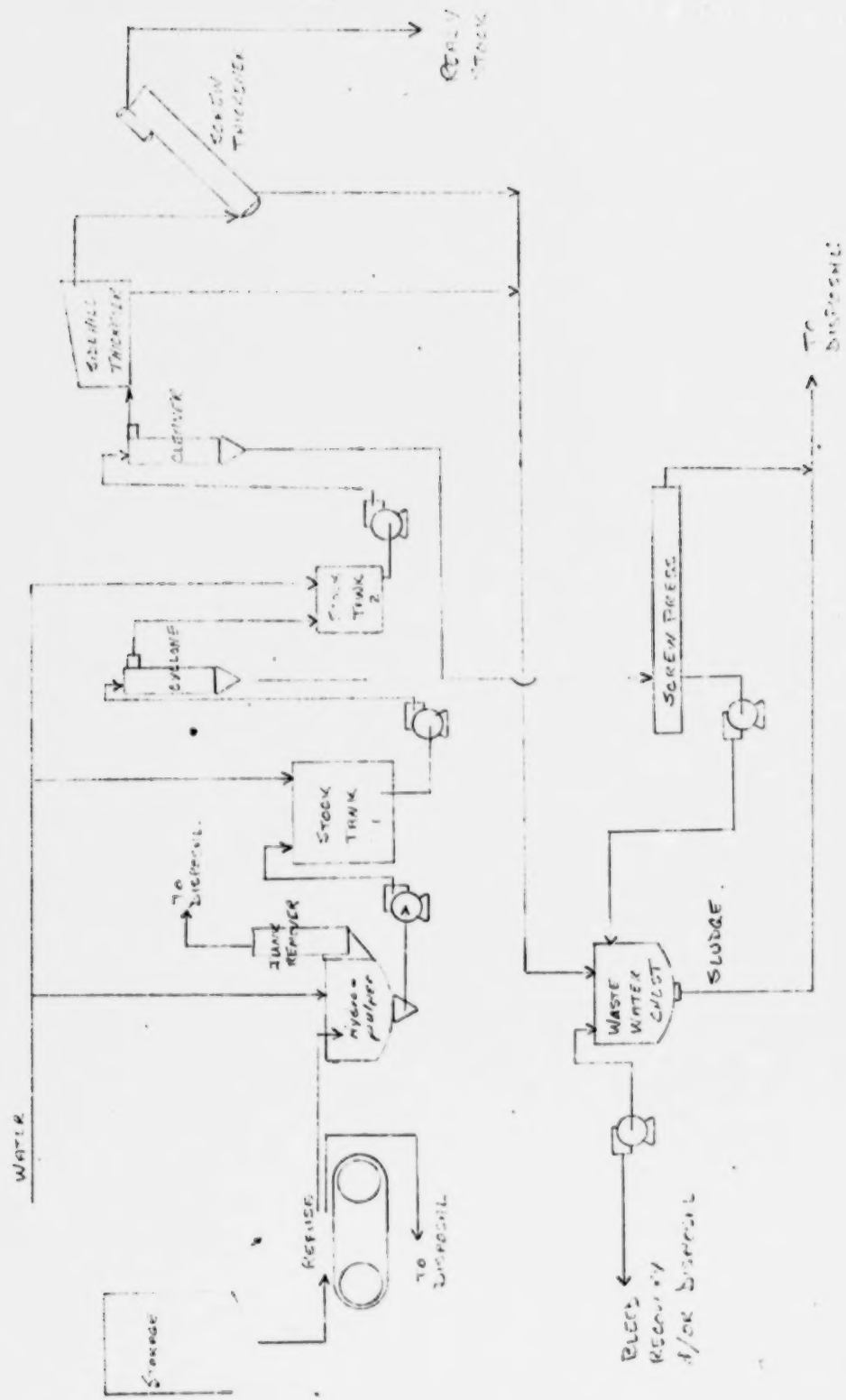
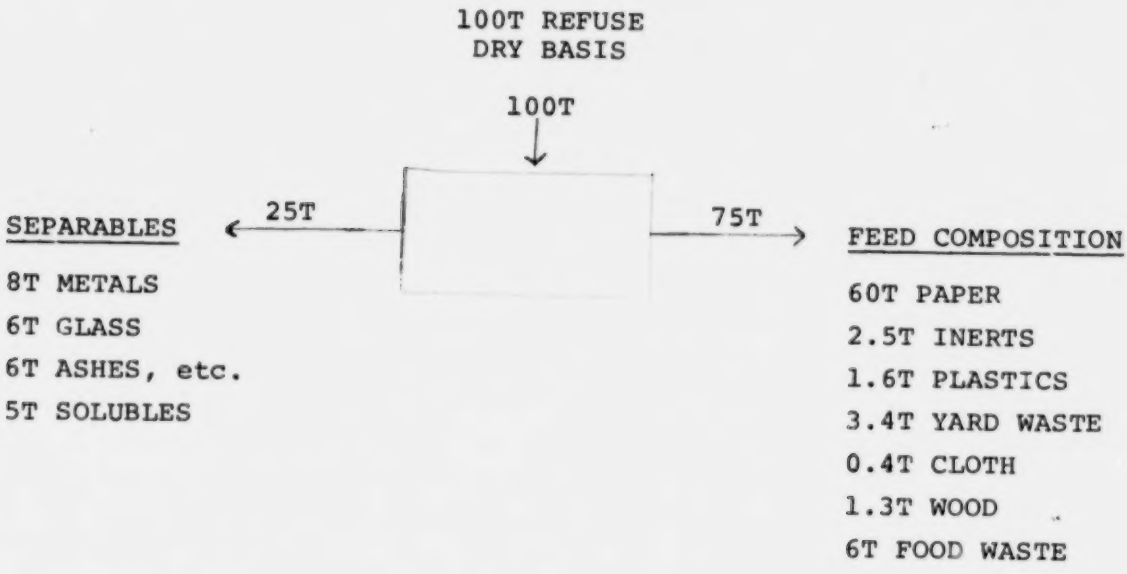


Figure 4-2

REFUSE STOCK PRETREATMENT FLOW SHEET

BASED ON REFUSE CONTAINING 60% PAPER



APPROXIMATE COMPOSITION OF FEED TO REACTOR

TOTAL CELLULOSE = 48T = 64% of FEED
SOLUBLE ORGANICS = 9.4T = 12.5% of FEED
SOLID INERTS = 17.6T = 23.5% of FEED

Figure 4-3
HYDRASPOSAL PROCESS

4.1.2 Hydrolysis and Concentration

The hydrolysis reactor and the concentration stage constitute the main body of the mass balance simulation program. The inputs to this stage of the process are the pretreated feedstock, lime, acid, and process water. The initial feed stream and the lime slurry both have a liquid content of 50%. 98% sulfuric acid is used as stock for the acid injection system.

The feed slurry is diluted to the proper solid/liquid ratio, and the mixture is heated by the direct injection of superheated steam at 500 psia. Sulfuric acid is then mixed with the slurry and the stream is screw-fed into the reactor. The reaction is quenched by flashing to 150°C, and the slurry is centrifuged to separate the solids. At this point, a fraction of the solids containing unreacted cellulose may be recycled to the slurry heater. The liquid stream, containing the sugar solution, is neutralized with the lime slurry and enters a second centrifuge where unreacted lime and precipitated calcium sulfate are removed. The sugar solution is finally concentrated in a series of long tube, vertical evaporators. The number of effects is calculated such that the concentration of the final sugar solution is 12 weight %. The long tube vertical evaporators are used because the experience with the bench scale reactor has shown that the effluent has the tendency to foam. The process flow sheet follows as Figure 4-4.

4.2 Costing Procedures

4.2.1 Assumptions

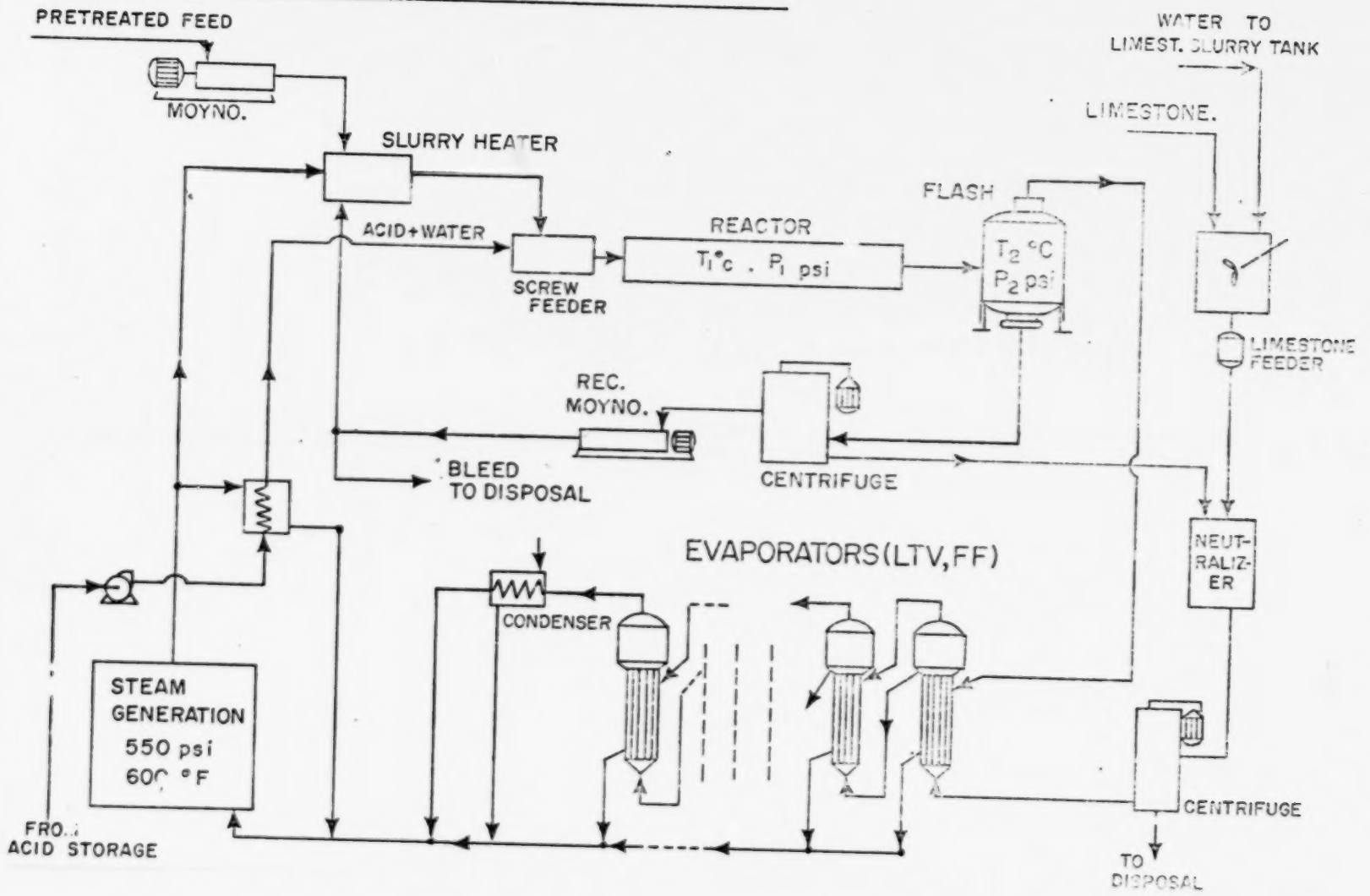
A number of simplifying assumptions have been made to facilitate the mass balance calculations. These assumptions are:

1) no reactions occur other than the hydrolysis reactions associated with the rate constants k_1 and k_2
2) liquid loss with the separated solids is 60% by weight of solids,
3) the weight of the decomposition products equals the weight of the sugar decomposed, and
4) all decomposed products are in liquid phase.

Data on the average refuse composition[7] indicates that most common urban refuse has a paper content within the range of 40

Figure 4-4

ACIDHYDROLYSIS PLANT FLOW SHEET



to 60%. In the cost analysis, therefore, two sample refuse sources were used as input for the mass balance simulation -- 40 and 60% paper content. The composition of the feedstock for the reactor stage is simplified to three major constituents, viz, cellulose, solid inerts and soluble organics. The composition of the feedstock for the two representative refuse samples is as follows:

	60% paper content	40% paper content
Cellulose	64%	51%
Solid inerts	23.5%	32%
Soluble organics	12.5%	17%

It is assumed that the paper contains 80 weight percent cellulose (amorphous and crystalline).

4.2.2 Capital Costing Procedures

Estimates for the initial capital investment were obtained using modular costing procedure as suggested by Guthrie[8]. Guthrie lists the purchased cost for each piece of equipment along with a scaling exponential factor and a direct material and labor factor. These factors are used to calculate the total installed equipment cost for the plant. The cost estimates were for the mid-1968; thus a 15% escalation was added to estimate the total installed equipment cost for January 1971. To obtain a value for the total capital investment, the remaining plant costs were determined as is shown in Table 4-1a.

4.2.3 Manufacturing Costs

The raw material demands for any given production capacity are based on the mass balance for the plant. The prices for the raw materials were obtained from the Oil, Paint and Drug Report[9]. The utility costs were estimated by the rates given by Peters and Timmerhaus[10]. The labor costs were estimated for three men operating the plant on a continuous 24-hour basis with one supervisor. The labor rates were taken from U.S. Department of Labor publications.

The direct charges such as fringe benefits, maintenance and repair, and plant supplies, as well as the indirect charges for payroll and plant overhead, are all calculated on percentage basis

Table 4-1a

CAPITAL COST ANALYSIS

Item	Estimation Procedure
Installed Cost	Total Purchase Cost (TPC) + Installation Fee
Freight and Taxes	TPC * 0.08
Engineering	(TPC + Field Mater.) .136
Land/Yard Improve.	(TPC) 0.19
Buildings	(TPC) 0.25
Piping	(TPC) 0.4
Services	(TPC) 0.1
Contingency and Fees	(Total Installation Cost) 0.18
Fixed Capital Investment (FCI)	Sum of Above
Working Capital	(FCI) .15
Total Capital Investment (TCI) with 15% Escal.	(TCI) 1.15

Table 4-1b

MANUFACTURING COST - Per Day Basis

Item	Estimation Procedure
Raw Material & Utilities	Fixed Rates
Labor	Fixed for 24 hr oper. 3 man
Super.	Fixed for 24 hr oper. 1 man
Fringe Benefits	(Labor + Super.) 0.15
Maintenance & Repair	(TCI) .05/365
Supplies	(50% M & R) .15
Taxes	(TCI) .02/365
Insurance	(TCI) .01/365
Payroll Overhead	.15 Total Labor & Main.
Lab Work	.15 Total Labor + Main.
Plant Overhead	.5 (Total Labor + Main.)
Residue Disp.	$2.5 per ton
Capital Recovery	$TCI/365\ [H_2\ (1+H_2)^{H_3}/(1+H_2)^{H_3} - 1]$ $H_2 = 0.4$ $H_3 = 20$ yrs.
Total	- - -

as shown in Table 4-1b. The fixed charges for the taxes and insurance are estimated as a percentage of the total capital investment. The charges for the initial capital investment are determined using a capital recovery factor which is based on the economic plant life of 20 years and a rate of return of 4%. Table 4-1b gives a complete listing of the manufacturing cost analysis.

4.2.4 Program Operation

The mass balance simulation program BALANCE was written by U. Karmarkar. The Dartmouth BASIC language was used for implementation on the Dartmouth Time Sharing System. BALANCE includes two sub-programs, FLOCOST and FLOBAL, which contain the printout routines for the cost analysis and the mass balance respectively. A complete listing of the main program and the two sub-programs is attached in Appendix I.

The mass balance simulation is organized on the basis of the process streams within the plant. The stream code numbers are shown on the simplified flow diagram of the hydrolysis plant in Figure 4-5. The equipment codes and the operating parameters are defined in Table 4-2. Table 4-3 lists a sample printout for a nominal 250-ton plant in which the process streams are identified only by numbers.

The external inputs which must be entered to run the program are given in detail with the program listing in Appendix II. These inputs include the total capacity of the plant on a per day basis, the total input of the pretreated feedstock to the reactor stage, the cost of pretreatment, and the municipal rates.

The design parameters for the plant operation are determined using an optimization routine in which the manufacturing cost is minimized. Optimal values are found for the variables acid concentration, recycle ratio, reaction temperature, residence time, and solid to liquid ratio. Constraints on these variables are set in lines 1450 through 1650. The limits on the acid concentration and the recycle ratio are set such that the numbers remain meaningful. For example, the recycle ratio can only vary between zero and one. The limits for the reaction temperature are set at

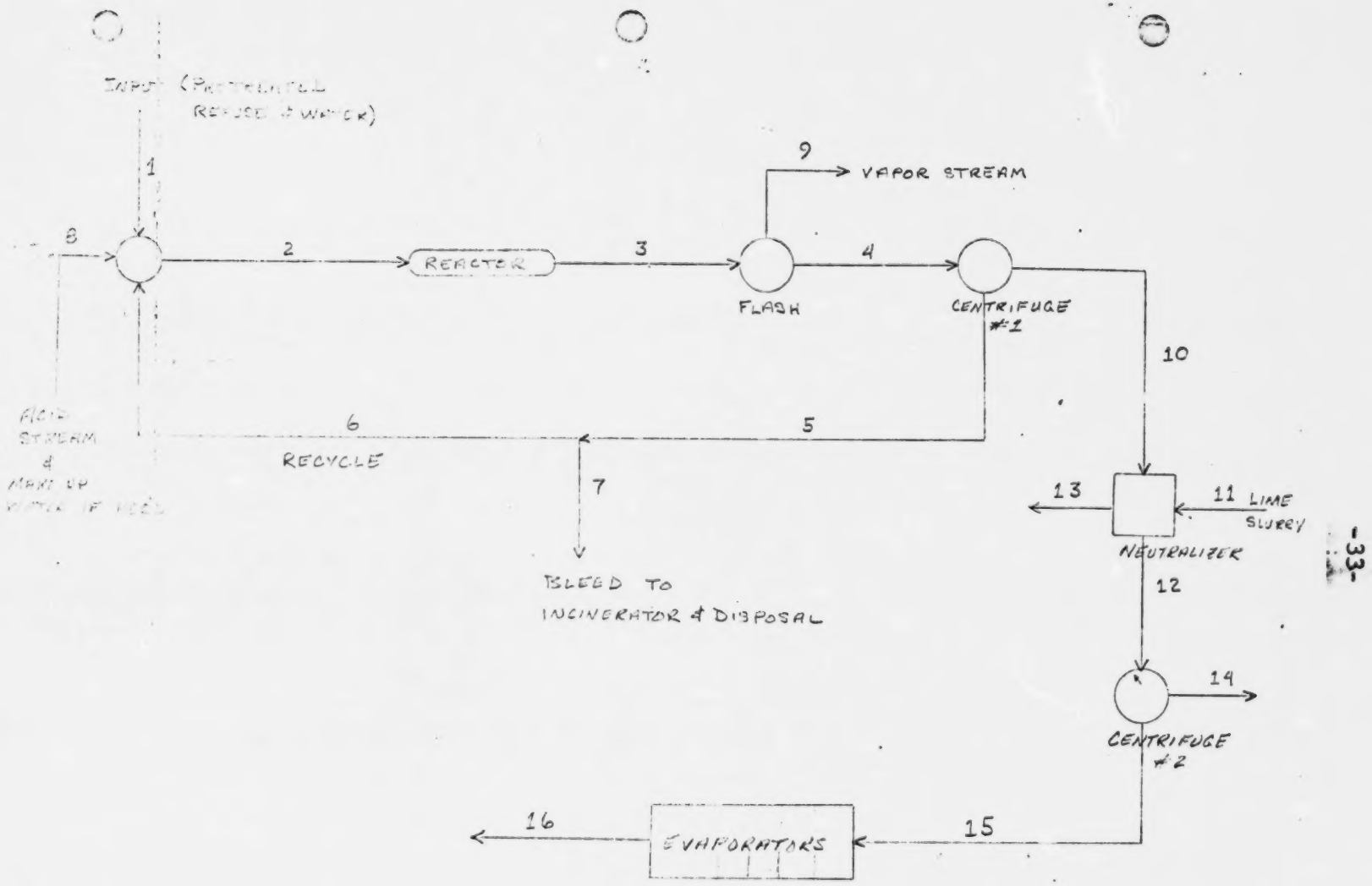


Figure 4-5

STREAM CODE NUMBERS FOR HYDROLYSIS PROCESS

Table 4-2

OPERATING PARAMETERS

Program Variable	Definition
R1	SOLID-TO-LIQUID RATIO
R2	ACID FRACTION IN LIQUID PHASE
R3	RECYCLE RATIO
T5	REACTOR RESIDENCE TIME (MIN.)
T1	REACTOR TEMP. (DEG. CENT.)
C8	FINAL SUGAR CONCENTRATION

EQUIPMENT CODE

Code Number	Equipment Description
1	SLURRY HEATER
2	REACTOR WITH SCREW FEEDER
3	FLASH
4	CENTRIFUGE #1
5	NEUTRALIZER
6	CENTRIFUGE #2
7	LIMESTONE STORAGE (10 DAYS)
8	LIME SLURRY TANK
9	ACID STORAGE
10	LIMESTONE CONVEYOR
11	OIL STORAGE
12	EVAPORATORS
13	STEAM GENERATION PLANT
14	MOYNO PUMP (TO REACTOR SYSTEM)
15	RECYCLE MOYNO PUMP
16	ACID FEED PUMP
17	WATER FEED PUMP
18	LIME SLURRY SCREW
19	ACID AND WATER HEATER
20	CONDENSER AND CONDENSER POT

MASS BALANCE FOR 250 TONS/DAY PLANT

Table 4-3

REFUSE INPUT 60% PAPER

R1 0.3 R2 0.019875
R3 0 T1 220.844
T5 0.25 C8 12

MIN COST = 1.65673 E-2

	LBS./HR.	TONS/HR	TONS/DAY	%COMP
STREAM 1				
CELL.	10000.	5.	120.	32.0963
WATER	15625	7.8125	187.5	50.1505
L.INERT	1843.75	0.921875	22.125	5.91775
S.INERTS	3687.5	1.84375	44.25	11.8355
TOT.MASS	31156.2	15.5781	373.875	
STREAM 2				
CELL.	10000.	5.	120.	14.8584
WATER	50741.9	25.3709	608.903	75.3942
ACID	1028.94	0.514472	12.3473	1.52885
L.INERT	1843.75	0.921875	22.125	2.73951
S.INERTS	3687.5	1.84375	44.25	5.47903
TOT.MASS	67302.1	33.651	807.625	
STREAM 3				
CELL.	1141.98	0.570991	13.7038	1.6968
WATER	49758.5	24.8793	597.102	73.9331
SUGAR	7174.08	3.58704	86.089	10.6595
ACID	1028.94	0.514472	12.3473	1.52885
L.INERT	1843.75	0.921875	22.125	2.73951
S.INERTS	3687.5	1.84375	44.25	5.47903
DECOMP	2667.31	1.33365	32.0077	3.96319
TOT.MASS	67302.1	33.651	807.625	
STREAM 4				
CELL.	1141.98	0.570991	13.7038	1.95478
WATER	41905.3	20.9526	502.863	71.7312
SUGAR	7174.08	3.58704	86.089	12.2802
L.INERT	1843.75	0.921875	22.125	3.15603
S.INERTS	3687.5	1.84375	44.25	6.31206
DECOMP	2667.31	1.33365	32.0077	4.56575
TOT.MASS	58419.9	29.2099	701.039	
STREAM 5				
CELL.	1141.98	0.570991	13.7038	14.7788
WATER	2223.18	1.11159	26.6781	28.7709
SUGAR	380.603	0.190301	4.56723	4.92551
ACID	54.588	0.027294	0.655056	0.706442
L.INERT	97.8154	4.89077 E-2	1.17379	1.26586
S.INERTS	3687.5	1.84375	44.25	47.7212
DECOMP	141.507	7.07536 E-2	1.69809	1.83129
TOT.MASS	7727.17	3.86359	92.7261	
STREAM 6				
TOT.MASS	0	0	0	

Table 4-3 (continued)

STREAM 7				
CELL.	1141.98	0.570991	13.7038	14.7788
WATER	2223.18	1.11159	26.6781	28.7709
SUGAR	380.603	0.190301	4.56723	4.92551
ACID	54.588	0.027294	0.655056	0.706442
L.INERT	97.8154	4.89077 E-2	1.17379	1.26586
S.INERTS	3687.5	1.84375	44.25	47.7212
DECOMP	141.507	7.07536 E-2	1.69809	1.83129
TOT.MASS	7727.17	3.86359	92.7261	
STREAM 8				
WATER	26167.7	13.0838	314.012	96.2166
ACID	1028.94	0.514472	12.3473	3.78336
STEAM	19700.6	9.85032	236.408	72.4379
TOT.MASS	27196.6	13.5983	326.359	
STREAM 9				
STEAM	7853.25	3.92662	94.239	
DIVISION BY ZERO IN 6246				
OVERFLOW IN 6246				
1.70141 E+38				
TOT.MASS	0	0	0	
STREAM 10				
WATER	39682.1	19.841	476.185	76.7224
SUGAR	6793.48	3.39674	81.5218	13.1347
ACID	974.357	0.487178	11.6923	1.88385
L.INERT	1745.93	0.872967	20.9512	3.37564
DECOMP	2525.8	1.2629	30.3096	4.88345
TOT.MASS	51721.7	25.8608	620.66	
STREAM 11				
WATER	1193.09	0.596545	14.3171	50
LIMEST.	1193.09	0.596545	14.3171	50
TOT.MASS	2386.18	1.19309	28.6342	
STREAM 12				
WATER	41054.1	20.5271	492.65	78.4701
SUGAR	6793.48	3.39674	81.5218	12.9849
L.INERT	1745.93	0.872967	20.9512	3.33715
DECOMP	2525.8	1.2629	30.3096	4.82777
LIMEST.	198.848	9.94241 E-2	2.38618	0.380075
TOT.MASS	52318.2	26.1591	627.818	
STREAM 13				
TOT.MASS	0	0	0	
STREAM 14				
WATER	733.037	0.366518	8.79644	64.9016
SUGAR	121.3	0.06065	1.4556	10.7397
L.INERT	31.1743	1.55871 E-2	0.374092	2.76011
DECOMP	45.0991	2.25495 E-2	0.541189	3.99298
LIMEST.	198.848	9.94241 E-2	2.38618	17.6056
TOT.MASS	1129.46	0.564729	13.5535	

Table 4-3 (continued)

STREAM 15				
WATER	40321.1	20.1606	483.853	78.7695
SUGAR	6672.18	3.33609	80.0662	13.0345
L.INERT	1714.76	0.85738	20.5771	3.34988
DECOMP	2480.7	1.24035	29.7684	4.84619
TOT.MASS	51188.7	25.5944	614.265	
STREAM 16				
SUGAR	6672.18	3.33609	80.0662	61.3949
L.INERT	1714.76	0.85738	20.5771	15.7786
DECOMP	2480.7	1.24035	29.7684	22.8265
TOT.MASS	10867.6	5.43382	130.412	
STREAM 17				
TOT.MASS	0	0	0	

CO2(13) 278.388 TOT.HEAT 7.93069 E+6

YIELD(4) 0.645723 YIELD(10) 0.611466

EQUIPMENT COSTS:
****

CODE	SIZE	P.C.	I.C.	%P.E.C.
1	73.8408	9805.49	24317.6	3.93992
2	4.22526	14168.2	42787.9	5.69289
3	33.3281	3369.62	13983.9	1.35394
4	107.593	60854.9	185608.	24.452
5	25.7176	3416.53	14178.6	1.37279
6	99.1576	57945.8	176735.	23.2831
7	4581.46	5540.93	15902.5	2.22639
8	38.1789	673.403	2794.62	0.270579
9	17614.2	4199.54	12052.7	1.68741
10	50	6357.71	19645.3	2.55458
11	37116.	4244.48	12393.9	1.70547
12	0	0	0	0
13	19700.6	54523.7	149940.	21.9081
14	18789.9	4596.64	15582.6	1.84697
15	0	0	0	0
16	417.093	650.645	2205.69	0.261434
17	17756.7	2678.05	9078.6	1.07606
18	217.106	271.157	919.222	0.108953
19	453.851	13763.3	44180.1	5.53019
20	2160.76	1815.04	5826.28	0.729298

TOTAL PURCHASED COST= 248875.
FIELD MATERIAL FACTOR= 0.995453
FIELD LABOUR FACTOR= 0.49705
MATERIAL+LABOUR FACTOR= 1.4925
DIRECT MATERIAL FACTOR= 1.99545
L/M FACTOR= 0.249091
INDIRECT FACTOR= 0.513551

Table 4-3 (continued)

CAPITAL COST ANALYSIS
****

ITEM	COST
INSTALLED COST	748132.
FREIGHT & TAXES	19910.
ENGINEERING	67540.1
LAND/YARD IMPROVEMENTS	47286.3
BUILDINGS	62218.8
PIPING	99550.
SERVICES	124438.
CONTINGENCY & FEES	134664.
FIXED CAPITAL INVESTMENT	1.30374 E+6
WORKING CAPITAL	195561.
TOTAL CAPITAL INVESTMENT	1.4993 E+6
T.C.I.WITH 15% ESCALATION	1.58926 E+6

F.C.I. FACTOR = 5.23852
T.C.I. FACTOR = 6.0243

MANUFACTURING COST ANALYSIS:(PER DAY BASIS)
****

ITEM	COST	CONTR.%
ACID	395.115	14.8934
LIMESTONE	186.122	7.01563
WATER	40.2095	1.51565
OIL	371.16	13.9904
POWER	25.3099	0.954025
LABOUR	216	8.14185
SUPERV.	84	3.16627
FRINGE BEN.	45.	1.69622
MAINT.& REPAIR	264.876	9.98417
SUPPLIES	19.8657	0.748813
TAXES	105.95	3.99367
INSURANCE	52.9752	1.99683
PAYROLL OVERHEAD	64.8657	2.44503
LAB WORK	64.8657	2.44503
PLANT OVERHEAD	216.219	8.1501
RESIDUE DISPOSAL	110.625	4.16987
CAPITAL RECOVERY	389.801	14.6931
TOTAL MANF.COST	2652.96	100

COST OF SUGAR= 1.65673 E-2 $/LB.SUGAR
COST OF PRETREATMENT= 7.02544 E-3 $/LB.SUGAR
EXPECTED MIN.SALES PRICE= 0.02 $/LB.SUGAR
EXPECTED MIN.NET PROFIT PER DAY=-575.314 $/DAY
MAX.BREAKEVEN DUMP FEE= 2.30125
MAX.NET PROFIT PER DAY= 1026.01 $/DAY
MIN.BREAKEVEN DUMP FEE=-4.10404

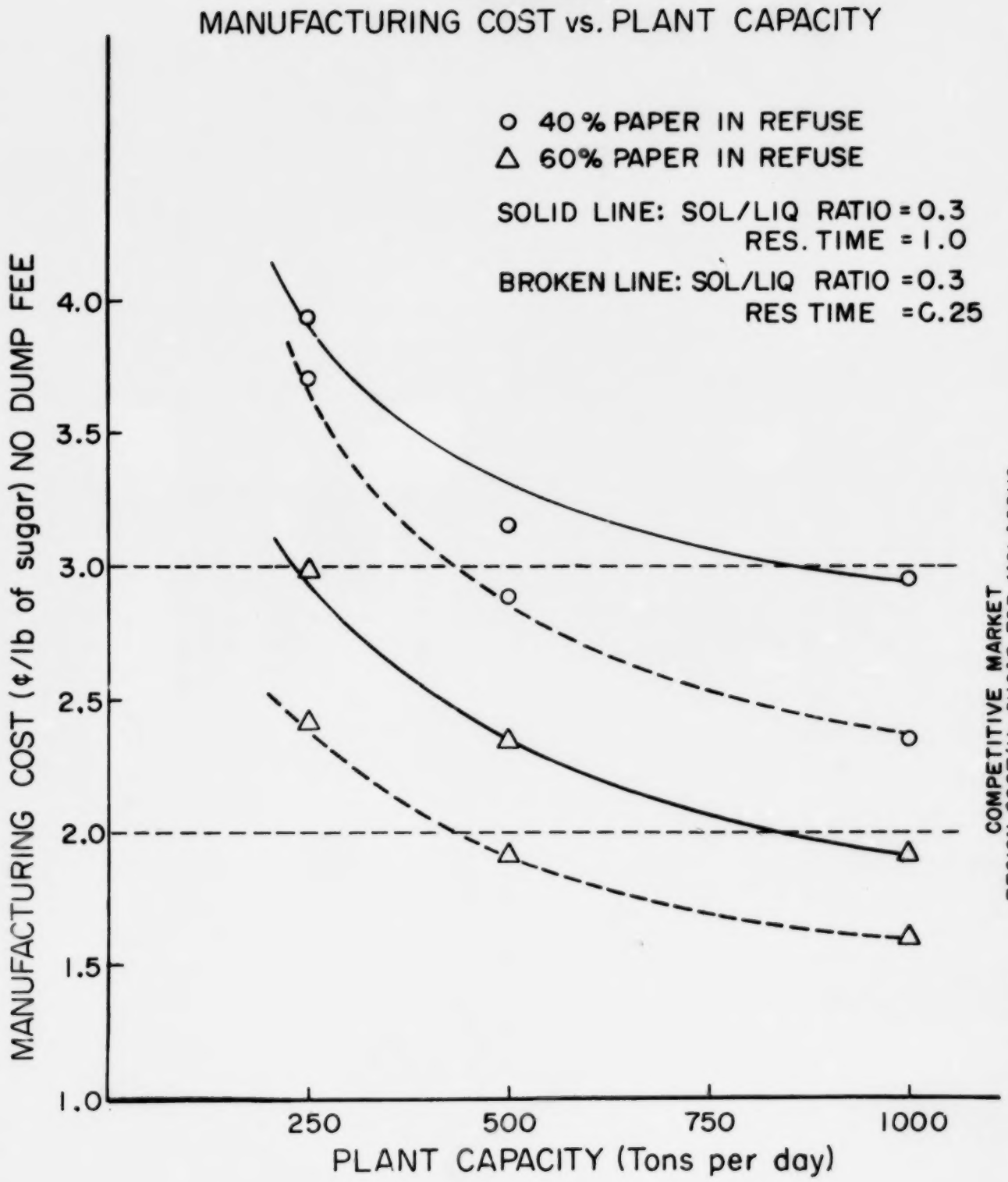
160 and 240°C. The optimal reaction time for temperatures above 240°C would be so short that imperfect mixing of the acid with the slurry would limit the yield. The constraints on the last two variables are determined by actual physical limitations of the process equipment. It should be noted from the data presented in Chapter 2 that much greater yields are theoretically possible by hydrolyzing at very high reaction temperatures, such as 240°C or above, and very short reaction times, such as less than 15 seconds. The actual constraints on the residence time are determined by the cost and the physical limitations of the process control system. In the present cost analysis, manufacturing cost of sugar is calculated for two levels of constraints on the minimum permissible residence time, viz, 15 seconds and 60 seconds. Increasing the solid to liquid ratio would directly increase the concentration of the product sugar solution. This relationship will be valid only as long as complete mixing of the reactants can be achieved. Data obtained from the Ionics Incorporated[11] have shown consistent hydrolysis results using a plug flow reactor system and a maximum solid to liquid ratio of 0.25. In the present cost analysis, two levels of constraints on the maximum permissible solid to liquid ratio are used, viz, 0.1 and 0.3. Effects of these variables and constraints on the manufacturing cost of sugar are given in the following section.

4.3 Cost Analysis

The mass balance simulation program previously described has been used to generate the information necessary for making a decision concerning the economic potential of the paper hydrolysis plant as a means of refuse disposal. For any fixed set of operational parameters, the manufacturing cost of sugar will be a function of the total refuse input and the refuse composition.

Figure 4-6 shows the effect of plant capacity (total raw waste input) on the manufacturing costs. Two refuse sources were studied; one containing 60% paper, the other containing 40% paper. Because the constraints on the operating parameters were known to have a significant effect on the manufacturing cost, two cases were also studied for each refuse source. The broken line

Figure 4-6


MANUFACTURING COST vs. PLANT CAPACITY
o 40 % PAPER IN REFUSE
Δ 60% PAPER IN REFUSE
SOLID LINE: SOL/LIQ RATIO = 0.3
RES. TIME = 1.0
BROKEN LINE: SOL/LIQ RATIO = 0.3
RES TIME = 0.25
MANUFACTURING COST (¢/lb of sugar) NO DUMP FEE
4.0
3.5
3.0
2.5
2.0
1.5
1.0
250
500
750
1000
PLANT CAPACITY (Tons per day)
COMPETITIVE MARKET
REGION COST/lb SUGAR FOR MOLASSUS

represents the most optimistic conditions with the solid/liquid ratio constrained to 0.3 and the minimum residence time constrained to 0.25 minutes. The solid line represents the less favorable limits of 1.0 minute for the residence time and 0.3 for the solid/liquid ratio. Figure 4-6 also shows the range of sugar cost per pound for blackstrap molasses as the raw material for a fermentation process. Blackstrap molasses is the most widely used fermentation raw material and must be considered the most competitive sugar containing product. Because it is an agricultural commodity, the selling price fluctuates a great deal. Data on the selling price of molasses was obtained from the Commodity Year 1968 Book. The molasses prices for 1966, 1967, and 1968 indicate a range of fluctuation between 2 cents and 3 cents per pound of sugar. This range is indicated as the competitive market region on Figure 4-6. Thus, plants which operate at a cost below that of molasses can be considered to be producing a salable raw material. It should be noted that the manufacturing costs given in Figure 4-6 do not include a dump fee.

The costs for incineration range from $7 per ton to $3 per ton of refuse input. A representative cost of $4 per ton was chosen with which to compare the hydrolysis process. Figure 4-7 shows the dump fees that would have to be charged in order to break even plotted against plant capacity. Given a fixed market price for sugar, the break-even dump fee is the charge per ton of refuse which would allow the plant to just break even or make a zero profit. A negative dump fee indicates the plant is producing sugar at a competitive cost. A dump fee less than $4 per ton of refuse allows the hydrolysis plant to compete with the incinerator process as a feasible alternative for refuse disposal. Here again, the two refuse sources were studied. The solid line represents the break-even dump fee which would be required to meet operating costs if the market price for sugar were 2¢ per pound; whereas, the data for the broken line was calculated on the basis of a 3¢ per pound market price. The critical plant parameter for all the points on this graph were 0.3 (solid/liquid ratio) and 1.0 minutes (residence time).

Figure 4-7

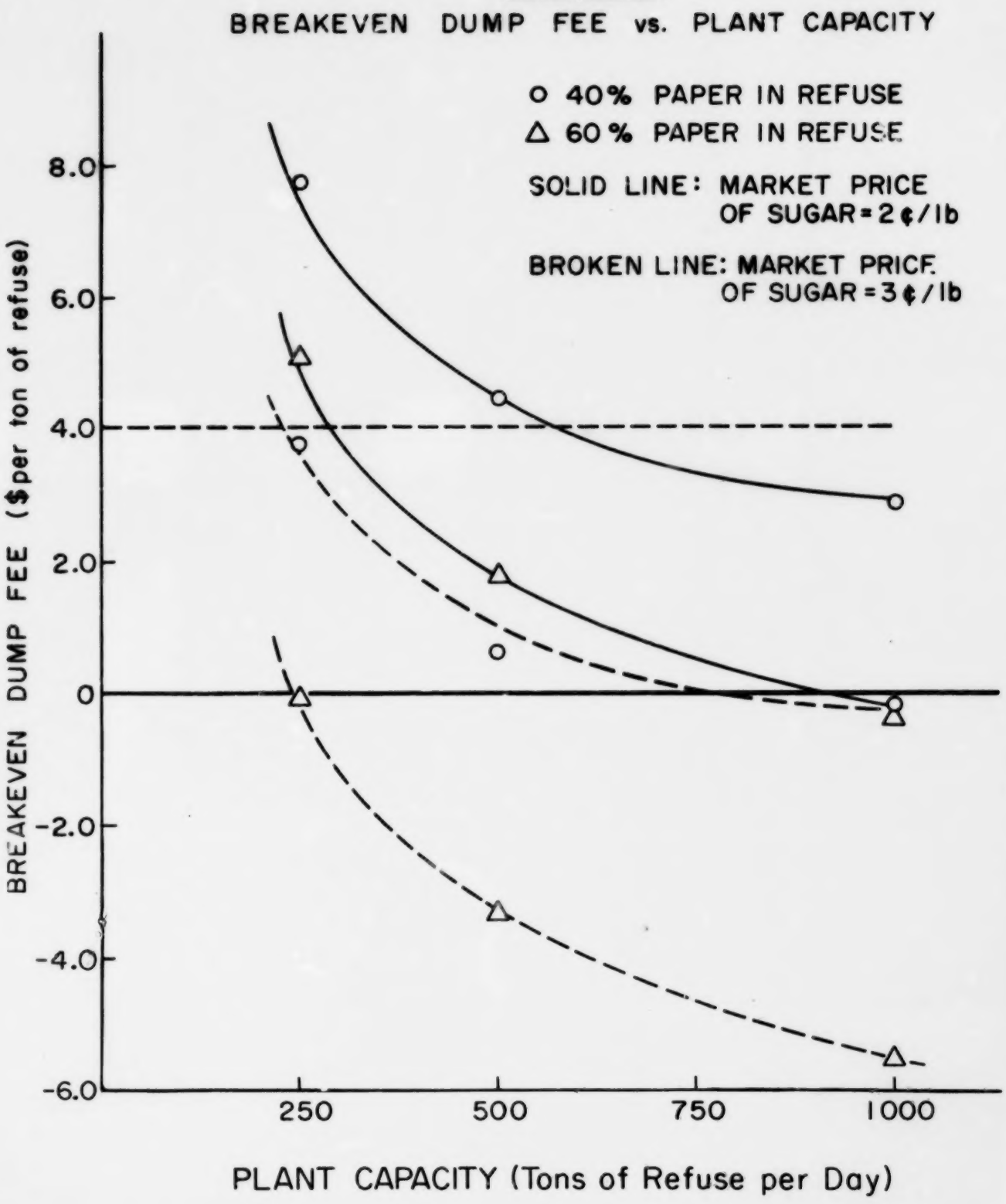

BREAKEVEN DUMP FEE vs. PLANT CAPACITY
o 40% PAPER IN REFUSE
Δ 60% PAPER IN REFUSE
SOLID LINE: MARKET PRICE OF SUGAR = 2¢/lb
BROKEN LINE: MARKET PRICE OF SUGAR = 3¢/lb
BREAKEVEN DUMP FEE ($per ton of refuse)
8.0
6.0
4.0
2.0
0
-2.0
-4.0
-6.0
250
500
750
1000
PLANT CAPACITY (Tons of Refuse per Day)

Figure 4-7 indicates that the competitive potential of the hydrolysis process drops markedly as the plant size decreases. This becomes critical for plant capacities under 500 tons per day if refuse with a 50% paper composition is used as input. For refuse containing 60% paper, the plant capacity can be as low as 300 tons per day and still remain feasible. In Figure 4-6, it is seen that a significant reduction of the manufacturing cost can be attained by improving the constraints on the operating parameters. In many cases, hydrolysis plants with a capacity below 500 tons per day could be operating under extremely marginal conditions; thus, a knowledge of the cost sensitivity to these critical plant parameters would be extremely advantageous. Table 4-4 lists the manufacturing cost of sugar as a function of the minimum residence time and the solid/liquid ratio. In the range studied, plant A represents the most ideal conditions with a solid/liquid ratio of 0.3 and residence time of 15 seconds, and plant I represents the worst case where the operating parameters are 0.1 and 60 seconds respectively. The cost estimates are calculated for a nominal 1000 ton plant, and the paper content of the refuse is 60%.

Tables 4-5, 4-6, and 4-7 list the manufacturing cost analyses for plants A, C, and G respectively. Comparing plants A and C, shows the cost benefit accrued by increasing the solid/liquid ratio for a constant residence time; whereas, a comparison of plant A and plant G isolates the effect of decreasing the minimum residence time. It can be concluded that the manufacturing cost is much more sensitive to changes in the solid/liquid ratio within the range studied. This is encouraging since solid/liquid ratios around 0.3 have been successfully used in an hydrolysis process by Ionics Incorporated[11]. Table 4-8 shows the major contribution to the daily manufacturing cost. The sensitive analysis in Table 4-4 accounts for the costs involved with the altered demands for oil and acid. Table 4-8 shows that, in the present cost analysis, the contributions due to the raw material requirements for fuel oil and acid dominate the total production cost. The contribution of the capital recovery cost is significantly smaller than the raw material requirements and it is unlikely that any increase due to an improved control system or reactor design would

TABLE 4.4

MANUFACTURING COST

RESIDENCE TIME (MIN)	Sol/Liq. Ratio 0.3	0.2	0.1
0.25	Plant A 1.61	Plant B 2.252	Plant C 3.626
0.5	Plant D 1.97	Plant E 2.28	Plant F 3.850
1.0	Plant G 1.93	Plant H 2.459	Plant I 4.128

The cost figures represent cents per pound of sugar produced. All the cost estimates are based on a plant capacity of 1,000 tons of refuse per day. The refuse contains 60% paper products.

Table 4-5

COST ANALYSIS FOR PLANT "A" -- REFUSE INPUT 60% PAPER

P.C. = Purchase Cost
I.C. = Installed Cost
%P.E.C. = % of Purchased Equipment Cost

MASS BALANCE FOR 1000 TONS/DAY PLANT

R1 0.3 R2 1.53047 E-2
R3 0 T1 225.008
T5 0.25 C8 12

MIN COST = 1.15128 E-2
YIELD(4) 0.607267 YIELD(10) 0.573247

EQUIPMENT COSTS:
****

CODE	SIZE	P.C.	I.C.	%P.E.C.
1	295.363	22527.1	55867.2	3.91342
2	16.9291	32582.3	98398.7	5.66022
3	133.739	8314.22	34504.	1.44435
4	429.148	139569.	425687.	24.2461
5	101.655	8347.74	34643.1	1.45017
6	389.124	131607.	401401.	22.8628
7	14067.5	8678.97	24908.7	1.50772
8	117.229	1320.09	5478.37	0.229327
9	54255.1	5885.39	16891.1	1.02241
10	50	6357.71	19645.3	1.10446
11	150618.	7432.77	21703.7	1.29122
12	0	0	0	0
13	81048.9	146748.	403557.	25.4931
14	81314.6	9846.63	33380.1	1.71056
15	0	0	0	0
16	1389.94	1511.05	5122.44	0.2625
17	76922.2	5739.82	19458.	0.997124
18	666.631	485.928	1647.29	8.44156 E-2
19	1847.51	34277.5	110031.	5.9547
20	8454.8	4405.65	14142.1	0.765352

TOTAL PURCHASED COST= 575637.
FIELD MATERIAL FACTOR= 0.996309
FIELD LABOUR FACTOR= 0.487045
MATERIAL+LABOUR FACTOR= 1.48335
DIRECT MATERIAL FACTOR= 1.99631
L/M FACTOR= 0.243973
INDIRECT FACTOR= 0.515872

Table 4-5 (continued)

CAPITAL COST ANALYSIS
****

ITEM	COST
INSTALLED COST	1.72647 E+6
FREIGHT & TAXES	46051.
ENGINEERING	156284.
LAND/YARD IMPROVEMENTS	109371.
BUILDINGS	143909.
PIPING	230255.
SERVICES	287819.
CONTINGENCY & FEES	310764.
FIXED CAPITAL INVESTMENT	3.01092 E+6
WORKING CAPITAL	451638.
TOTAL CAPITAL INVESTMENT	3.46256 E+6
T.C.I.WITH 15% ESCALATION	3.98194 E+6

F.C.I. FACTOR = 5.23058
T.C.I. FACTOR = 6.01517

MANUFACTURING COST ANALYSIS:(PER DAY BASIS)
****

ITEM	COST	CONTR.%
ACID	1217.03	17.5342
LIMESTONE	571.493	8.23372
WATER	162.122	2.33575
OIL	1506.18	21.7001
POWER	102.815	1.48129
LABOUR	216	3.112
SUPERV.	84	1.21022
FRINGE BEN.	45.	0.648332
MAINT.& REPAIR	663.657	9.56156
SUPPLIES	49.7742	0.717117
TAXES	265.463	3.82462
INSURANCE	132.731	1.91231
PAYROLL OVERHEAD	94.7742	1.36545
LAB WORK	94.7742	1.36545
PLANT OVERHEAD	315.914	4.5515
RESIDUE DISPOSAL	442.5	6.37527
CAPITAL RECOVERY	976.66	14.0711
TOTAL MANF.COST	6940.88	100

COST OF SUGAR= 1.15128 E-2 $/LB.SUGAR
COST OF PRETREATMENT= 4.56142 E-3 $/LB.SUGAR
EXPECTED MIN.SALES PRICE= 0.02 $/LB.SUGAR
EXPECTED MIN.NET PROFIT PER DAY= 2366.75 $/DAY
MAX.BREAKEVEN DUMP FEE=-2.36675
MAX.NET PROFIT PER DAY= 8395.57 $/DAY
MIN.BREAKEVEN DUMP FEE=-8.39557

Table 4-6

COST ANALYSIS FOR PLANT "C"

REFUSE INPUT 60% PAPER

MASS BALANCE FOR 1000 TONS/DAY PLANT

R1 0.1 R2 1.48281 E-2
R3 0 T1 225.234
T5 0.25 C8 12

MIN COST = 3.18241 E-2
YIELD(4) 0.602608 YIELD(10) 0.589543

EVAPORATOR DATA:

EFFECT	TEMP	PRESS	VAP	AREA
1	275.144	58.9078	50834.4	2702.11
2	264.737	49.2358	56004.5	11553.8
3	253.906	39.5638	60666.	14288.3
4	242.639	29.8919	64727.9	17835.5
5	230.925	20.2199	68101.3	22756.4

STEAM REQUIREMENT= 35636.2 LBS/HR
ENTERING SUGAR CONCN.= 4.68227 E-2 LEAVING CONC.= 12

EQUIPMENT COSTS:
****

CODE	SIZE	P.C.	I.C.	%P.E.C.
1	295.363	22527.1	55867.2	1.64076
2	44.3611	58077.	175393.	4.23004
3	353.2	15630.3	64865.8	1.13843
4	1119.36	248089.	756673.	18.0696
5	287.914	16423.1	68156.1	1.19618
6	1103.87	246024.	750373.	17.9192
7	42375.9	13490.5	38717.8	0.982584
8	353.133	2558.26	10616.8	0.186331
9	157697.	8105.69	23263.3	0.590378
10	50	5357.71	19645.3	0.463064
11	701551	13753.8	40161.2	1.00176
12	12410.6	253982.	838140.	18.4988
13	255686.	327978.	901938.	23.8883
14	81660.3	9868.38	33453.8	0.718764
15	0	0	0	0
16	4057.14	3198.42	10842.6	0.232957
17	377048.	13118.3	44471.1	0.955473
18	2008.11	862.185	2922.81	6.27973 E-2
19	8983.66	95823.6	307594.	6.97932
20	68101.3	17097.9	54884.4	1.24533

TOTAL PURCHASED COST= 1.37296 E+6
FIELD MATERIAL FACTOR= 1.02415
FIELD LABOUR FACTOR= 0.535702
MATERIAL+LABOUR FACTOR= 1.55985
DIRECT MATERIAL FACTOR= 2.02415
L/P FACTOR= 0.264655
INDIRECT FACTOR= 0.497749

Table 4-6 (continued)

CAPITAL COST ANALYSIS
****

ITEM	COST
INSTALLED COST	4.19798 E+6
FREIGHT & TAXES	109837.
ENGINEERING	377956.
LAND/YARD IMPROVEMENTS	260863.
BUILDINGS	343241.
PIPING	549186.
SERVICES	686482.
CONTINGENCY & FEES	755636
FIXED CAPITAL INVESTMENT	7.28118 E+6
WORKING CAPITAL	1.09218 E+6
TOTAL CAPITAL INVESTMENT	8.37336 E+6
T.C.I.WITH 15% ESCALATION	9.62936 E+6

F.C.I. FACTOR = 5.30325
T.C.I. FACTOR = 6.09874

MANUFACTURING COST ANALYSIS:(PER DAY BASIS)
****

ITEM	COST	CONTR.%
ACID	3537.4	17.9537
LIMESTONE	1721.52	8.73741
WATER	427.883	2.17168
OIL	7015.51	35.6065
POWER	281.114	1.42677
LABOUR	216	1.09629
SUPERV.	84	0.426333
FRINGE BEN.	45.	0.228393
MAINT.& REPAIR	1604.89	8.14547
SUPPLIES	120.367	0.61091
TAXES	641.957	3.25819
INSURANCE	320.979	1.62909
PAYROLL OVERHEAD	165.367	0.839303
LAB WORK	165.367	0.839303
PLANT OVERHEAD	551.223	2.79768
RESIDUE DISPOSAL	442.5	2.24586
CAPITAL RECOVERY	2361.82	11.9872
TOTAL MANF.COST	19702.9	100

COST OF SUGAR= 3.18241 E-2 $/LB.SUGAR
COST OF PRETREATMENT= 4.4418 E-3 $/LB.SUGAR
EXPECTED MIN.SALES PRICE= 0.02 $/LB.SUGAR
EXPECTED MIN.NET PROFIT PER DAY=-10070.5 $/DAY
MAX.BREAKEVEN DUMP FEE= 10.0705
MAX.NET PROFIT PER DAY=-3879.34 $/DAY
MIN.BREAKEVEN DUMP FEE= 3.87934

Table 4-7

COST ANALYSIS FOR PLANT "G" -- REFUSE INPUT 60% PAPER

MASS BALANCE FOR 1000 TONS/DAY PLANT

R1 0.3 R2 1.45002 E-2
R3 0 T1 209.904
T5 1 C8 12

MIN COST = 1.39571 E-2
YIELD(4) 0.518056 YIELD(10) 0.487406

EVAPORATOR DATA:

EFFECT	TEMP	PRESS	VAP	AREA
1	256.267	41.6399	10560.2	31.7829
2	224.034	14.7	16875.6	1541.55

STEAM REQUIREMENT= 559.973 LBS/HR
ENTERING SUGAR CONCN.= 0.103705 LEAVING CONC.= 12

EQUIPMENT COSTS:
****

CODE	SIZE	P.C.	I.C.	%P.E.C.
1	295.363	22527.1	55867.2	3.54999
2	67.736	74867.7	226100.	11.7982
3	134.266	8335.52	34592.4	1.31357
4	444.614	142566.	434826.	22.4666
5	104.541	8501.04	35279.3	1.33966
6	399.347	133671.	407696.	21.0648
7	13283.8	8482.22	24344.	1.33669
8	110.698	1275.46	5293.14	0.200996
9	51403.3	5790.82	16619.7	0.912561
10	50	6357.71	19645.3	1.00189
11	160480.	7623.74	22261.3	1.20141
12	786.665	21133.9	69741.9	3.33044
13	74214.5	137972.	379423.	21.7427
14	60695.2	8457.47	28670.8	1.33279
15	0	0	0	0
16	982.95	1185.64	4019.32	0.186842
17	59575.1	5025.57	17036.7	0.791966
18	629.49	471.656	1598.91	7.43271 E2
19	1776.88	33420.	107278.	5.26657
20	16875.6	6904.15	22162.3	1.08801

TOTAL PURCHASED COST= 634568.
FIELD MATERIAL FACTOR= 0.976336
FIELD LABOUR FACTOR= 0.506577
MATERIAL+LABOUR FACTOR= 1.48291
DIRECT MATERIAL FACTOR= 1.97634
L/M FACTOR= 0.256321
INDIRECT FACTOR= 0.530877

Table 4-7 (continued)

CAPITAL COST ANALYSIS
****

ITEM	COST
INSTALLED COST	1.91246 E+6
FREIGHT & TAXES	50765.5
ENGINEERING	170560.
LAND/YARD IMPROVEMENTS	120568.
BUILDINGS	158642.
PIPING	253827.
SERVICES	317284.
CONTINGENCY & FEES	344242.
FIXED CAPITAL INVESTMENT	3.32835 E+6
WORKING CAPITAL	499252.
TOTAL CAPITAL INVESTMENT	3.8276 E+6
T.C.I.WITH 15% ESCALATION	4.40174 E+6

F.C.I. FACTOR = 5.24505
T.C.I. FACTOR = 6.03181

MANUFACTURING COST ANALYSIS:(PER DAY BASIS)
****

ITEM	COST	CONTR.%
ACID	1153.06	16.0998
LIMESTONE	539.653	7.53501
WATER	142.348	1.98756
OIL	1604.8	22.4073
POWER	92.6873	1.29417
LABOUR	216	3.01594
SUPERV.	84	1.17287
FRINGE BEN.	45.	0.628322
MAINT & REPAIR	733.623	10.2434
SUPPLIES	55.0217	0.768252
TAXES	293.449	4.09734
INSURANCE	146.725	2.04867
PAYROLL OVERHEAD	100.022	1.39657
LAB WORK	100.022	1.39657
PLANT OVERHEAD	333.406	4.65525
RESIDUE DISPOSAL	442.5	6.1785
CAPITAL RECOVERY	1079.62	15.0745
TOTAL MANF.COST	7161.94	100

COST OF SUGAR= 1.39571 E-2 $/LB.SUGAR
COST OF PRETREATMENT= 5.35917 E-3 $/LB.SUGAR
EXPECTED MIN.SALES PRICE= 0.02 $/LB.SUGAR
EXPECTED MIN.NET PROFIT PER DAY= 350.852 $/DAY
MAX.BREAKEVEN DUMP FEE=-0.350852
MAX.NET PROFIT PER DAY= 5482.25 $/DAY
MIN.BREAKEVEN DUMP FEE=-5.48225

have an overriding effect on the total manufacturing cost. It is apparent, however, that more data on these effects must be obtained before a decision can be made concerning the cost benefit which results from changes in the critical plant operating parameters.

TABLE 4.8

% Contribution to Tot. Manufacturing Cost

	Cap. Recov. Cost	Oil	Acid
Plant G	15%	22%	16%
Plant C	12%	35%	18%

The results of this cost analysis have demonstrated the desirability of the hydrolysis process as an economically favorable alternative means of refuse disposal for larger population centers. It is concluded that the manufacturing cost of sugar will be competitive for plants treating more than 500 tons of waste per day and the refuse contains at least 40% paper. This capacity corresponds to a population source of 200,000. Hydrolysis plants with a capacity below 500 tons per day can be considered if the limits on the sol/liq ratio (0.3) and the minimum residence time (25 min.) in the reactor stage are obtainable.

5. FERMENTATION OF SUGARS OBTAINED FROM HYDROLYSIS OF REFUSE

5.1 Introduction:

The sugar from hydrolysis of the cellulose in refuse in turn can be used to produce ethanol by fermentation. When Fagan (3) studied the kinetics of refuse hydrolysis with sulfuric acid, his cost analysis of the conversion of the sugar by fermentation showed that ethanol can be produced at 52¢/gallons, which is its current selling price, and thus offset some of the refuse disposal cost to the community undertaking such a venture. The objectives of this chapter are (I) to study the fermentation kinetics and yield of ethanol from hydrolyzed solid wastes, (II) to investigate the effect of the source of solid wastes on the yield of ethanol, (III) to analyze the quality of effluent from such a process and the type of treatment required to bring the quality of the effluent to acceptable levels, and (IV) to estimate the economic feasibility of such a project.

5.2 Materials and Methods:

In order to study the fermentation of sugar from hydrolysis of refuse, samples of this type of material had to be prepared. As reported in Chapter 2, the kinetics of the acid hydrolysis of cellulose follows Saemen's model[2] for a first order consecutive reaction. The cellulose hydrolyzes to sugar which in turn decomposes. Since the activation energy for the hydrolysis step is greater than the activation energy for the decomposition step, increasing the reaction temperature increases the rate of the first step over the second. As a result, the highest yield of sugar occurs at high temperatures and short reaction times. For example, an isothermal reactor at 230°C with 1% acid concentration converts 51% of the potential sugar (cellulose content) in solid waste to sugar in 20 seconds. While on a large scale the easiest way to realize such a short reaction time at 230°C is with a steady state plug reactor, on a laboratory scale it was convenient to prepare the sugar solution from hydrolyzed refuse in a batch reactor. Naturally the batch reactor with a long heat up time will not give the maximum sugar yield of an isothermal reactor but it will provide

material containing the representative impurities that will be present in the sugar solution.

There are two types of impurities present in the hydrolyzed solid wastes which may interfere with the subsequent fermentation. Some impurities come with the raw material (solid wastes) such as all the non-cellulosic portion of the refuse.

The second type of impurities are those formed during the hydrolysis reaction, such as the decomposed sugars. Since the hydrolysis was carried out in a non-isothermal batch reactor, the proportion of the impurities of the second type is increased relative to the amount of sugar in the hydrolyzate when compared to an isothermal plug flow reactor. It was felt that fermentation studies will be most useful on material which is indicative of the sugar solution obtainable from a large scale plug flow reactor. Therefore, the solution provided from the batch reactor was supplemented with pure sugar addition to compensate for the sugar lost due to the longer reaction time.

The reactor used for hydrolysis was cylindrical steel bomb (I.D. = 5 cm, Height = 13 cm), made of Carpenter 20 steel, and fitted with a thermocouple well. About 40 g of solid waste were hydrolyzed per batch, using 200 ml of 1 weight % sulfuric acid. The hydrolysis was carried out non-isothermally by placing the reactor in a constant temperature oil bath. The reactor was removed from the oil bath when the temperature of the reactants reached 230°C and quenched under tap water. The hydrolyzed refuse was filtered and neutralized with calcium carbonate. The precipitate ($CaSO_4$) was removed by filtration. Because of the inhibiting concentration of calcium sulfate in the solution, this sugar solution did not yield ethanol when fermented. Therefore the sugar solution was heated to 100°C for fifteen minutes to precipitate more completely the calcium sulfate from the solution which was removed by a second filtration. Supplementary sugar (dextrose) was added to this solution for reasons mentioned earlier. The resulting solution was then fermented.

Fermentation was carried out in a cylindrical glass vessel

(I.D.=10 cm, Height=20 cm). The working volume of the fermenter was 800 ml. For every run, 800 ml of hydrolyzed refuse prepared as described earlier, was introduced into the fermenter. Generally, this solution contained 5% sugar, about 8% decomposed sugar and other unknown impurities formed by the reaction of the sulfuric acid with the components present in the refuse. For run numbers 9 and 10, the initial sugar concentration was about 12% and the initial concentration of the decomposed sugar was about 24%. The contents of the fermenter were stirred by a magnetic stirrer. The temperature of the fermenter was regulated to ±0.5°C with a temperature controller. The pH of the medium was adjusted by adding 2.5 N solution of sodium carbonate every hour. Microorganisms (Saccharomyces Cerevisiae) were grown separately and injected into the fermenter after the desired conditions of temperature and the pH of the medium were attained. For run numbers 1-9, the initial cell concentration was between 45 and 50 millions per ml of the solution. For run numbers 10 and 11, the initial cell concentration was about 10 million per ml.

The first sample from the fermenter was drawn 30 minutes after the microorganisms were innoculated. Subsequent samples were drawn at an interval of two hours. Each sample was analyzed for sugar concentration, ethanol concentration, and cell concentration. Sugar is analyzed colorometrically[12] by reacting hexose sugars with O-Toluidine to obtain a green colored complex. The sample was analyzed for cell concentration by staining the cells with Magdala Red (a product of Aniline and Chemical Company, New York). This stain contains Tetrazolium ion which combines with the active enzyme to form a colored complex. The total number of viable cells is then counted in a hemacytometer. The ethanol content of the solution was estimated[13] by oxidizing the ethanol to acetaldehyde when then reacts with sodium nitroprusside to form a blue complex. The concentration of ethanol was then estimated by using a spectrophotometer.

Run numbers 1 to 6 were carried out using ground solid waste sample obtained from the Bureau of Solid Wastes Management, Cincinnati, Ohio. This was a typical city refuse sample of unknown general composition which contained 2º weight % cellulose.

Run number 2 is a replicate of run number 1. A two level factorial design was carried out in run numbers 3 to 6 to study the effect of variation in the operating temperature and the pH on the yield of ethanol and the rate of fermentation. Two levels of temperature -- 30 and 35°C -- and two levels of pH -- 4.4 and 5.0 were used in these four runs. The yield of ethanol was not signigicantly affected by either of these variables in the above mentioned ranges. As can be seen from the time required to complete the fermentation in Table 5-1, a higher rate of fermentation was obtained at a temperature of 30°C than at 35°C, and at a pH of 4.4 than at pH of 5.0.

Run 7 was carried out using a sample of household refuse. This sample contained food wastes (e.g., rice, tomato soup, meat pieces, potato peels, stale lettuce leaves), paper and plastic packaging materials, beer cans, cigarette butts and ashes, a copy of the New York Times, and a garbage bag. 200 g of this sample containing 79 g of cellulosic material, 28 g of plastics, 87 g of food wastes, and 6 g of metals, was mixed thoroughly in a Waring blender and hydrolyzed in a batch reactor. Though this sample contained 43.5% food wastes as compared with 12% food wastes obtained in a typical municipal refuse[7], the yield of ethanol and the rate of fermentation were not significantly affected by this change in the composition of the refuse.

The last set of four experiments (runs 8 - 11) were done on refuse samples obtained from the Black Clawson "FIBRECLAIM" process. These four experiments were a two-level factorial design to study the effect of variation in the initial concentrations of sugar and microorganisms. Two levels of initial sugar concentration -- 4% and 12% -- and two levels of initial microorganisms concentration -- 10 and 50 million/ml -- were used. The yield of ethanol was not significantly affected by a variation in either of these variables. The rate of fermentation was higher for the high initial cell concentration. Also the fermentation rate with high initial sugar concentration was greater than with the low initial sugar concentration, as can be seen from the results of runs 8 to 11 in Table 5-1. For a microorganism concentration of about 49 million/ml, it took 12 hours to ferment 4% sugar solution

in run 8, whereas it took less than double the time (20 hours) to ferment three times as much sugar (12.6%) in run 9. Finally, comparing the yields of runs 8 - 11 with the yields obtained in other runs, one observes that the yield is not affected by a change in the source of the refuse. The results of all the runs are given in Appendix III. The summary of these results is given in Table 5-1.

5.3 Analysis of Experimental Data

Luedeking and Piret[14] proposed a model

$$\frac{dP}{dt} = k_1 \cdot \frac{dm}{dt} + k_2 \cdot m \qquad (1)$$

to account for the behavior of batch lactic acid fermentation; where

P = concentration of product
m = concentration of microorganisms
t = time of fermentation
k_1 and k_2 are constants

Aiyer and Luedeking[15] showed that this model was applicable to the batch fermentation of dextrose to ethanol. Experimental data obtained in this study were used to fit the above model. Integrating equation (1) gives

$$P_{t_i} - P_{t_1} = k_1 (m_{t_i} - m_1) + k_2 \int_{t_1}^{t_1} m\,dt \qquad (2)$$

A polynomial of the form

$$m = \sum_{1=0}^{4} A_i t^{2i} \qquad (3)$$

was fitted to the data of cell concentration (m) as a function of time. Using equations (2) and (3), k_1 and k_2 were estimated by the method of least squares. The fit of the model and the significance of each of the parameters was tested by the analysis of variance technique[16]. The summary of parameter estimation is given in Table 5-2. In all except one run, the only statistically significant parameter is k_1 which indicates that the rate of ethanol formation is proportional to the growth rate of microorganisms.

Table 5-1 Summary of Experiments

Run No.	Operating Conditions		Initial Conc.		Run Time	Yield of Alcohol = α	Increase in yeast during the Run	Remarks
	Temp. (°C)	pH	Sugar (g/100 ml)	Yeast (millions/ml)	(Hours)	$\frac{\text{Exptl.}}{\text{theor.}} \times 100$	(millions/ml)	
1	35	5.0	4.50	50.0	11.0	91.2	4.0	Ground refuse from P.H.S.
2	35	5.0	4.20	49.5	11.0	88.1	3.8	"
3	35	5.0	3.90	46.7	10.75	93.9	3.3	"
4	30	5.0	4.25	48.2	10.0	87.6	11.5	"
5	30	4.4	4.10	48.45	9.0	89.5	6.15	"
6	35	4.4	3.95	45.3	10.0	85.0	6.3	"
7	30	4.4	4.15	48.2	10.0	93.0	11.1	Household refuse (High % of organics)
8	35	5.0	4.10	49.1	12	91.5	4.5	Black Clawson Sample
9	35	5.0	12.6	47.4	20	90.0	8.6	"
10	35	5.0	11.8	9.2	32	91.6	6.6	"
11	35	5.0	4.05	9.8	26	86.5	4.3	"

Table 5-2 Summary of Analysis of Variance

Run No.	Value of Parameters k_1 g/hr. ethanol/ growth rate of microorganisms (millions/ml/hour)	Value of Parameters k_2 g/hr. ethanol/ cell concentration (millions/ ml)	Fit of the Model (95% Level of confidence)	Significance of Parameters k_1	Significance of Parameters k_2	Mean Absolute deviation (Exptl. Ethanol - Predicted) g/100 ml.
1	0.558	2.48 E-4	Model Fits	Significant	Not significant	0.072
2	0.533	-1.60 E-4	"	"	"	0.105
3	0.588	1.35 E-4	"	"	"	0.060
4	0.154	-5.85 E-4	"	"	"	0.046
5	0.339	6.00 E-4	"	"	Significant	0.023
6	0.288	7.00 E-6	"	"	Not significant	0.052
7	0.228	-5.05 E-4	"	"	"	0.099
8	0.377	-3.68 E-4	"	"	"	0.118
9	0.681	-2.4 E-4	"	"	"	0.095
10	0.795	-9.9 E-4	"	"	"	0.058
11	0.396	-2.1 E-4	"	"	"	0.042

Sugar concentration at any time was predicted by a yield equation of the form

$$S_{t_1} - S_{t_i} = \frac{180}{92} \cdot \alpha \cdot (P_{t_i} - P_1)$$

where S_{t_1} = initial sugar concentration in g/100 ml.

S_{t_i} = sugar concentration at time t_i in g/100 ml.

P_{t_1} = initial ethanol concentration in g/100 ml.

P_{t_i} = ethanol concentration at time t_i in g/100 ml.

α = percent ethanol yield ($\frac{\text{experimental yield}}{\text{theoretical yield}}$ x 100)

Values of α are given in Table 5-1. The factor 180/92 is the stoichiometric ratio between sugar consumed and ethanol formed. Fitted curves of the cell concentration and the predicted values of ethanol and sugar concentrations are given in Appendix IV.

As will be discussed in the cost analysis section, the yield of ethanol in fermentation is the most important parameter in determining the optimum operating conditions. From Table 5-1 it can be seen that there is no significant effect of temperature, pH, initial sugar concentration, and the initial microorganism concentration on the yield of ethanol in the factor space explored. For a given yield of ethanol, it is desirable to have a higher rate of fermentation. From equation (1), the rate of product formation is equal to k_1 times the growth rate of microorganisms. Run numbers 8 to 11 show that it is desirable to operate with high initial microorganism concentration and an initial sugar concentration of about 12 g/100 ml, because these conditions give a higher rate of fermentation for the same yield. Effect of variation in the temperature and pH on the rate of fermentation are not very pronounced in the factor space examined, but a temperature of 30°C and a pH of 4.4 give a slightly higher rate of fermentation, as can be seen from the time required to complete the fermentation for runs 1 to 6 in Table 5-1.

5.4 Effluent from the Process

Effluent from the fermenter in run number 6 was analyzed

for the biological oxygen demand (B.O.D.). The ethanol in the liquor was stripped away by distillation prior to the test. The standard method for the determination of the BOD in the waste liquor had to be modified due to the presence of ferrous ions and nitrite ions. Ferrous ions are present in the waste liquor because of the iron content in the solid wastes. Nitrite ions are formed by the biological degradation of proteinaceous materials during fermentation. Hence the BOD was determined using the Rideal-Stewart modification of the Winkler (Azide) procedure[17]. The five day BOD of the unfiltered liquor sample was 6,550 ppm and that of the filtered sample was 6,070 ppm.

It is difficult to estimate the cost of effluent treatment for this process since ethanol production by fermentation is a dying art and the lack of knowledge about the difficulties that may be encountered while treating the effluent generated by acid hydrolysis followed by fermentation of the solid wastes. Examining the effluent treatment costs in industries like foods and paper[18], and taking into account the high BOD of the effluent in this process, it is estimated that the capital investment in the treatment plant will be $12,000 per 25,000 gallons of effluent treated. This process will be producing 400,000 gallons per day of effluent for a 250 tons/day refuse processing capacity. Hence the capital investment for the effluent treatment plant will be $192,000. Assuming the economic life of the plant to be 20 years[18] and an interest rate of 4%, the cost of effluent treatment works out to be 0.7¢/gal. of ethanol produced.

5.5 Ethanol Manufacturing Cost

Fagan[3] has estimated the cost of ethanol manufactur using sugar obtained from the acid hydrolysis of refuse. His analysis is summarized in Table 5-3. From the Table, it can be seen that the cost of sugar accounts for over 83% of the total cost. The experimental work carried out on the fermentation of hydrolyzed solid wastes in this work shows that the yield of ethanol varies from 85% to 93.9%, as compared to the yield of 80% used in Fagan's cost estimate. It should be noted that the entire factor space of operating and process variables was not exhaustively searched for the maximum yield. Fagan[3] has pointed out the difficulty

Table 5-3 Analysis of Ethanol Manufacturing Cost

Source	Cost ¢/gallon	% of total cost
1) Cost of Sugar @ 2.8¢/lb.	43.0	83.25
2) Raw materials (excluding sugar, labor and utilities)	2.2	4.25
3) Maintenance, overheads and fixed costs	6.5	12.50
TOTAL COST	51.7	100.00

Basis:

a) Sugars obtained from a refuse processing plant with a capacity of 250 T/day of solid refuse.

b) Paper content of the refuse = 50%.

c) No fee is charged to the community to get rid of its refuse.

d) Initial capital investment = $\$1.7 \times 10^6$.

in estimating the initial capital investment in ethanol fermentation plants, mainly because no new plants have been built recently to manufacture ethanol by fermentation. Karandikar[19] has estimated an initial capital investment of 2.1 million for such a process. Table 5-4 shows the sensitivity of the cost of ethanol manufacture using two parameters -- initial capital investment and ethanol yield in fermentation. Three levels of initial capital investment are used -- $1.7 million, as used by Fagan[3], $2.1 million as estimated by Karandikar[19], and $2.55 million an estimate 50% higher than Fagan's estimate. Three levels of ethanol yields are used -- the maximum yield (93.9%), average yield (90%), and the minimum yield (85%) obtained in this work. It should be noted that in the above cost estimation, no dumping fee is charged to the community to get rid of its refuse, and a fee of $3/ton is charged to the entrepreneur to get rid of the process waste. The current market price for ethanol is 52¢/gal.

5.6 Discussion

In a typical ethanol manufacture from fermentation of molasses, the initial sugar concentration is 10-13 g/100 ml, the time of fermentation varies between 30 to 50 hours, and the average ethanol yield is 80%[20]. The time of fermentation for run number 9, with initial sugar concentration of 12.6 g/100 ml was 20 hours with a 90% yield. The yield of ethanol in all the eleven runs varied from 85% to 93.9% -- average yield being 90%. One possible explanation for higher yields and faster rate of fermentation is that all the impurities present in the refuse and formed during hydrolysis, form an environment conducive to the formation of ethanol from the hydrolyzed refuse using Saccharomyces Cerevisiae.

Cost analysis of ethanol manufacture shows that the cost of sugar accounts for 83% of the total manufacturing cost. Hence, the yield of ethanol in fermentation is the most important parameter in examining the effect of the operating variables. The first set of six experiments (run No. 1 to 6) shows that there is no significant change in the yield of ethanol, either due to change in temperature between 30°C and 35°C, or due to a change

Table 5-4 Manufacturing Cost of Ethanol

Cost of Ethanol ¢/gallon

yield	85%	90%	93.9%
Initial Investment			
$1.7 million	49.1	46.8	45.3
$2.1 million	50.7	48.4	46.9
$2.55 million	52.35	50.05	48.55

Note: The plant charges no fee to process the community's refuse.

in pH between 4.4 and 5.0. The insensitivity of the ethanol yield to the temperature and pH changes in the ranges mentioned above should make it easy to control the fermentation process in a full size plant. The second set of four experiments (run No. 8 to 11) shows that though the yield of ethanol is not significantly affected by the changes in the initial sugar concentration between 4 g/100 ml and 12 g/100 ml or the initial cell concentration between 10 million/ml and 50 million/ml, it is preferable to operate at high initial cell concentration and an initial sugar concentration of 12 g/100 ml mainly because of faster rate of fermentation. The source of solid wastes also does not appreciably affect the yield as is seen from the results of run numbers 1-6 (using ground solid wastes from the city of Cincinnati), run number 7 (using a typical household garbage sample) and run numbers 8-11, which used samples obtained from the Black Clawson "FIBRECLAIM" process. This is an encouraging evidence in view of the fact that in an actual process using solid wastes, the composition of the raw material may vary from day to day.

Analysis of experimental data shows that the rate of ethanol formation is directly proportional to the growth rate of microorganisms and it does not depend upon the concentration of microorganisms. Examination of the values of k_2 in all the eleven runs shows that they lie on either side of the value zero, from $6 * 10^{-4}$ to $-9.9 * 10^{-4}$, indicating that it is due to a noise in the experimental data, and k_2 is not a significant parameter.

Cost analysis of ethanol manufacturing shows that except for the worst possible case (the lowest ethanol yield of 85% and the maximum initial investment of $2.55 million), the cost of ethanol manufacturing compares favorably with the current selling price of ethanol which is 52¢/gal. A community of 200,000 people paying $3/ton to get rid of its solid wastes, spends about $600,000/year as a cost of refuse disposal. If this community decides to invest $2.1 million for the proposed method of refuse disposal, the process will operate at nearly break-even point, thus helping the community to get rid of its wastes at no additional cost. This would mean an indirect savings of $600,000 per year to the community.

6. CONCLUSIONS AND EXTENSIONS

Batch kinetic study of the acid hydrolysis of cellulose found in paper showed that the hydrolysis reaction could be modeled by the Cellulose $\longrightarrow$ Sugar $\longrightarrow$ Decomposed sugar irreversible reaction, as found by Saeman[2]. The sugar yield depends on the time of reaction, temperature and acid concentration in the manner shown in Figure 2-3. Kinetic study of the acid hydrolysis of ground solid refuse in a continuous isothermal plug flow reactor showed that operating at high temperatures and low reaction times gives yields consistent with yields predicted by the batch study.

Economic analysis of the acid hydrolysis process presented in Chapter 4 shows that sugar can be produced by this process at a price comparable to the existing market price for sugar from molasses, which is in the range of \$0.02 - .03 per pound. Process economics depends greatly upon operating the process at high solid to liquid ratio. The present kinetic study -- both batch and continuous -- used a solid to liquid ratio of .01, whereas in the economic analysis, a solid to liquid ratio of .1 to .3 is used.

Since the parameter solid to liquid ratio influences the economics of this process so much, it is essential to carry out further work to verify the effects of high solid to liquid ratios on the kinetics of hydrolysis in a flow reactor. In such a study, it is important to check upon (i) whether mass transfer of acid becomes the rate controlling step in the kinetics of hydrolysis of refuse, (ii) whether the slurry can be heated to the reaction temperature in a short time to avoid thermal decomposition of cellulose, and (iii) whether such a slurry can be pumped easily through the process.

It was pointed out in Chapter 3 that the operating difficulties experienced in the bench scale flow reactor could be minimized if the scale of operation is increased. Such a flow reactor would be able to generate more reliable data on kinetics of hydrolysis and facilitate the study of the effects of acid concentration and solid to liquid ratio on the sugar yields. Moreover, the concentration of the sugar solution from the hydrolysis reaction is increased as the solid/liquid is increased. Consequently, the size of the evaporator, and in turn, the cost of manufacture,

decrease with increasing solid/liquid ratio.

Fermentation of sugars obtained from the acid hydrolysis of refuse shows that ethanol yields in the range of 85% to 93.9% are obtainable, as compared with a yield of 80% obtained in the industrial fermentation of molasses to ethanol. Kinetics of fermentation shows that the rate of ethanol formation is directly proportional to the growth rate of microorganisms. Economic analysis of the fermentation process shows that ethanol can be produced at a price lower than 52¢ per gallon, which is the current market price for ethanol, and thus offset some of the refuse disposal cost of a community undertaking such a venture.

An empirical relation of the growth rate of microorganisms was used to fit the kinetic data. It is recommended that in future extension of this work, the growth rate should be related to the environmental conditions such as temperature, pH and substrate concentration. Such a functional relationship will aid in optimizing the operating and process variables in the fermentation unit so as to minimize the manufacturing cost of ethanol.

The effluent from the fermentation process has a B.O.D. of about 6,000 ppm. The cost of effluent treatment, estimated at 0.7¢ per gallon of ethanol produced, could be reduced considerably if useful by-products, mainly amyl alcohols, could be recovered from the effluent. This will reduce the B.O.D. of the effluent and increase the revenue earned through the sale of by-products. Since the effluent in this process is obtained from the acid hydrolysis followed by fermentation of solid wastes, effects of all the impurities present in the effluent on the by-product recovery process will have to be studied.

The quantity of solid waste generated being so high, alternate uses of sugar obtained from the acid hydrolysis of refuse should be studied. Conversion of sugar to amino acids like Lysine, or to the production of single cell proteins or as a direct feed, are some of the alternate uses which should be looked into.

REFERENCES

1. Porteous, A., "A Profitable Means of Municipal Refuse Disposal" Thayer School of Engineering, Dartmouth College, Hanover, N.H., November 1966.

2. Saeman, J., "Kinetics of Wood Saccharification", Ind. Eng. Chem., January 1945.

3. Fagan, R.D., "The Acid Hydrolysis of Refuse", M.E. Thesis, Thayer School of Engineering, Dartmouth College, Hanover, N.H., July 1969.

4. Browning, B.L., The Chemistry of Wood, Interscience Publishers, New York, 1963

5. Wilde and Beightler, Foundations of Optimization, Prentice-Hall, 1967

6. Black Clawson Co., Research Center Technical Report, "Fibreclaim System", Middletown, Ohio, and private telephone conversations.

7. "Comprehensive Studies of Solid Waste Management", Third Annual Report, Report No. 70-2, University of California, Berkeley, June 1970.

8. Gutherie, K.M., "Capital Cost Estimating", Chemical Engineering, March 24, 1969, pp. 114-142.

9. Oil, Paint and Drug Report, June 16, 1969

10. Peters and Timmerhaus, Plant Design and Economics for Chemical Engineering, McGraw Hill, New York, N.Y., 1968

11. McNulty, K.V., Conversion of Organic Solid Wastes into Yeast, Research Division, Ionics, Inc.; Contract No. PH 86-67-204, February 11, 1968.

12. Nature, Vol. 183, p. 108, 1959.

13. Emmanel, Pfeil and Goldback, H., Kliniche Wochenschrift, "Eine Einfache Katalytische Methode zur Athylakkohol-bestimmung", Vol. 35, pp. 191-2, 1957.

14. Luedeking and Piret, Biotech. and Bioeng., Vol. 1, 393, 1959.

15. Aiyer and Luedeking, Chem. Eng. Progress Symposium Series, No. 55, 1966.

16. Draper and Smith, Applied Regression Analysis, John Wiley, New York, 1966.

17. Theroux, F.R., Eldrige, E.F., and Mallmann, W., Laboratory Manual for Chemical and Bacterial Analysis of Water and Sewage, McGraw Hill, New York, 1943

18. U.S. Federal Water Pollution Control Administration, "The Cost of Clean Water", Vol. III.

19. Karandikar, S., "Fermentation of Hydrolyzed Solid Wastes", M.S. Thesis, Thayer School of Engineering, Dartmouth College, Hanover, N.H., June 1971.

20. Shreve, R.N., Chemical Process Industries, McGraw Hill, New York, 1956.

Appendix I-1 Listing of Executive Program BALANCE


```
BALANCE

100 'ACID HYDROLYSIS PROCESS
110 '**********************
120 '
130 'THIS PROGRAM CONSISTS OF A MASS BALANCE SIMULATION PROGRAM,
140 'A COSTING ROUTINE, AND A SEARCH ROUTINE BASED ON THE DTSS
150 'LIBRARY PROGRAM " MINMUM***"(SOURCE: PROF.A.O.CONVERSE)
160 'THE SEARCH ROUTINE IS USED TO FIND THE PLANT PARAMETERS THAT
170 'MINIMISE THE COST OF MANUFACTURING THE PRODUCT SUGAR SOLUTION.
180 '
190 '
200 '
210 'INITIALISATION :
220 SUB FLOBAL;FLOCOST
230 DIM W(20),O(20),A(20),D(20),S(20),I(20),C(20),X(20),Y(20),Z(20)
240 DIM N(20),L(20),Q(20),B(20),K(20),F(20),G(20),H(20),V(20),E(20)
250 DIM T(6)
260 DIM P(6,2),M(7,7),R(7,7),J(7,1),U(7,1)
265 DIM A$(17)
270 '
280 READ G9,W9    'WT. OF O.DRY FEED,WT. OF WATER IN FEED STRM(TONS/DAY)
290 DATA 750,750
300 READ M1,M2,M3    'MOL.WTS.OF CELL.,SUGAR,WATER
310 DATA 162.14,180.14,18
320 READ P1,P2,P3   'WT.FRNS.OF CELL.,S.INERTS,& SOL.ORGS.IN FEED.
330 DATA .64,.236,.118
340 READ T2    'TEMP IN FLASH(DEG.CENT.)
350 DATA 150
360 READ A1,N1,E1,A2,N2,E2   'ARRHENIUS PARAMETERS
370 DATA 28.2105,1.7808,45127,4.90486,0.555121,32800
380 READ H2,H3    'INTEREST RATE,YEARS OF OPERATION
390 DATA .04,20
395 LET C3 = 2.75
400 LET J7=0
410 LET N=6
420 LET W2 = 1000
430 FOR I=1 TO N
440 READ P(I,1)
450 NEXT I
460 FOR I=1 TO N
470 READ T(I)
480 NEXT I
510 READ M
520 DATA 0.04,.7,1
530 DATA 230 'REACTOR TEMP(DEG.C.)
540 DATA .1    'SOL/LIQ RATIO
550 DATA 12    'OUTLET SUGAR CONCN.
560 DATA .001,.01,.1,1,.05,1.0
570 DATA 30
580 '
590 LET N=5
```

BALANCE (CONTINUED)

```
600 '"MINMUM" - FUNCTION MINIMISING ROUTINE, USING THE SECTIONING
610 'ALGORITHM DESCRIBED IN "OPTIMIZATION",BY PROF.A.O.CONVERSE
620 GOSUB 1330
630 LET F4=F
650 FOR N5=1 TO N
660 LET P(N5,2)=P(N5,1)
670 NEXT N5
680 FOR N6=1 TO M
690 FOR N5=1 TO N
700 FOR N4=2 TO 3
710 LET P(N5,1)=P(N5,1)+ (-1)↑N4*T(N5)
720 GOSUB 1330
730 IF F>=F4 THEN 760
740 LET P(N5,2)=P(N5,1)
750 GO TO 850
760 LET P(N5,1)=P(N5,2)
770 NEXT N4
780 NEXT N5
790 FOR N5=1 TO N
800 LET T(N5)=T(N5)/2
810 NEXT N5
820 NEXT N6
830 LET T=F4
840 GO TO 990
850 LET M5=0
860 LET F4=F
870 LET P(N5,1)=P(N5,1)+(-1)↑N4*T(N5)
880 GOSUB 1330
890 IF F<F4 THEN 920
900 LET P(N5,1)=P(N5,2)
910 GO TO 780
920 LET M5=M5+1
930 LET P(N5,2)=P(N5,1)
940 IF M5<3 THEN 860
960 LET T(N5)=T(N5)*2
980 GO TO 860
990 LET J9=1
1000 PRINT "THE MIN.COST OF SUGAR IS:";F;"$/LB."
1005 PRINT "Z3";
1006 INPUT Z3
1010 PRINT
1020 GOSUB 1330
1030 'REACTOR SECTION MASS BALANCE
1040 '
1050 'FUNCTION DEFINITIONS :
1060 '----------------------
1070 'PRESSURE (PSIA) AS A FN.OF TEMP.(DEG.F)--RANGE(50-600 DEG.F)
1080 DEF FNP(X)
1090 LET FNP=2.3395-8.81123E-2*X+1.17997E-3*X↑2-6.55206E-6*X↑3
1100 LET FNP=FNP+1.99202E-8*X↑4
```

BALANCE (CONTINUED)

```
1110 FNEND
1120 'LIQ.ENTHALPY (BTU/LB) / TEMP(DEG.F)--RANGE(50-600 DEG.F)
1130 DEF FNL(X)
1140 LET FNL=-30.8986+0.978717*X+1.09489E-4*X↑2-3.64545E-7*X↑3
1150 LET FNL=FNL+7.58916E-10*X↑4
1160 FNEND
1170 'VAPOUR ENTHAL(BTU/LB) / TEMP(DEG.F)--RANGE(50-600 DEG.F)
1180 DEF FNS(X)
1190 LET FNS=1060.45+0.480283*X-3.69591E-4*X↑2+8.07117E-7*X↑3
1200 LET FNS=FNS-1.75319E-9*X↑4
1210 FNEND
1220 'TEMP(DEG.F) / PRESSURE(PSIA)--RANGE(200-600 DEG.F)
1230 DEF FNT(X)
1240 LET FNT=204.937+1.33679*X-2.59432E-3*X↑2+2.30354E-6*X↑3
1250 LET FNT=FNT-6.9772E-10*X↑4
1260 FNEND
1270 'VOLUME OF STREAM (X): X=1 TO 10
1280 DEF FNV(Y)
1290 LET X=Y
1300 LET FNV=(W(X)+A(X)/1.8+C(X)/1.3+S(X)/1.6+O(X)/1.2+I(X)/1.3+D(X)/1.1)
1310 LET FNV=FNV/62.5
1320 FNEND
1330 'MASS BALANCE INITIALISATION :
1340 IF J7=1 THEN 1380
1350 LET G9 = G9*2000/24
1360 LET W9 = W9*2000/24
1370 LET J7=1
1380 LET R1=P(5,1)    'SOL/LIQ RATIO
1390 LET R2=P(1,1)    'ACID FR.IN LIQ PHASE
1400 LET R3=P(2,1)    'RECYCLE RATIO
1410 LET T5=P(3,1)    'REACTOR RES.TIME(MINS)
1420 LET T1=P(4,1)    'REACTOR TEMP(DEG.C.)
1430 LET C8=P(6,1)    'OUTLET SUGAR CONCN.
1440 '
1450 'CONSTRAINTS ON PROCESS PARAMETERS FOLLOW :
1460 IF R1<.3 THEN 1480
1470 LET R1=.3
1480 IF R1>0 THEN 1500
1490 LET R1=1E-3
1500 IF R2<.9999 THEN 1520
1510 LET R2=.999
1520 IF R2>0 THEN 1540
1530 LET R2=1E-5
1540 IF R3<.9999 THEN 1560
1550 LET R3=.999
1560 IF R3>=0 THEN 1580
1570 LET R3=0
1580 IF T5<100 THEN 1600
1590 LET T5 = 100
1600 IF T5>.25 THEN 1620
```

BALANCE (CONTINUED)

```
1610 LET T5=.25
1620 IF T1<240 THEN 1640
1630 LET T1=240
1640 IF T1>160 THEN 1660
1650 LET T1=160
1660 IF C8<15 GO TO 1680
1670 LET C8=15
1680 IF C8>3.9 THEN 1700
1690 LET C8=3.9
1700 '
1710 'MASS BALANCE CALCULATIONS
1720 'STREAM 1 :
1730 LET C(1)=G9*P1
1740 LET I(1)=G9*P2
1750 LET O(1)=G9*P3
1760 LET W(1)=W9
1770 '
1780 'STREAM 2:
1790 LET R4=(R3/(1-R3))              '(RECYCLE/BLEED)
1800 LET C(2)=C(1)+C(6)
1810 LET I(2)=I(1)+R4*I(1)
1820 LET O(2)=O(1)+R4*O(1)
1830 LET S(2)=S(6)
1840 LET D(2)=D(6)
1850 LET L(2)=(C(2)+O(1)+I(2))/R1
1860 LET A(2)=R2*L(2)
1870 LET W(2)=L(2)-R4*O(1)-A(2)-D(2)-S(2)
1880 '
1890 'REACTOR :
1900 LET K1=A1*(R2*100)↑N1*EXP(-E1/(1.98*(T1+273)))*1E15
1910 LET K2=A2*(R2*100)↑N2*EXP(-E2/(1.98*(T1+273)))*1E14
1920 LET F1=EXP(-K1*T5)              'MOL.FRN.CELL.UNCONVERTED
1930 LET F2=(K1/(K2-K1))*(EXP(-K1*T5)-EXP(-K2*T5))  'MOL.FRN.SUG.YIELD
1940 LET F3=EXP(-K2*T5)              'MOL.FRN.ENTERING SUG. UNCONVERTED
1950 '
1960 'STREAM 3 :
1970 LET C(3)=C(2)*F1
1980 LET I(3)=I(2)
1990 LET A(3)=A(2)
2000 LET O(3)=O(2)
2010 LET S(3)=C(2)*F2*(M2/M1)
2020 LET S(3)=S(3)+S(2)*F3
2030 LET D(3)=C(2)*(1-F1-F2)*(M2/M1)
2040 LET D(3)=D(3)+D(2)+S(2)*(1-F3)
2050 LET W(3)=W(2)+C(2)-C(3)-S(3)-D(3)
2060 LET L(3)=O(3)+S(3)+W(3)+D(3)+A(3)
2070 '
2080 'FLASH & STREAM 4 :
2090 LET T8=(T1*9/5)+32
2100 LET T9=(T2*9/5)+32
```

BALANCE (CONTINUED)

```
2110 LET V(9)=(W(3)+A(3)*.4+(S(3)+I(3)+C(3)+D(3)+O(3))*.35)*(T8-T9)
2120 LET V(9)=V(9)/(FNS(T9)-FNL(T9))
2130 LET W(4)=W(3)-V(9)
2140 LET L(4)=W(4)+A(3)+S(3)+O(3)+D(3)
2150 LET S(4)=S(3)
2160 LET D(4)=D(3)
2170 LET O(4)=O(3)
2180 LET I(4)=I(3)
2190 LET C(4)=C(3)
2200 '
2210 'STREAM 5 :
2220 LET I(5)=I(3)
2230 LET C(5)=C(3)
2240 LET L(5)=0.6*(C(5)+I(5))
2250 LET LS=L(5)/L(4)
2260 LET W(5)=W(4)*LS
2270 LET S(5)=S(3)*LS
2280 LET O(5)=O(3)*LS
2290 LET D(5)=D(3)*LS
2300 LET A(5)=A(3)*LS
2310 '
2320 'STREAM 6:
2330 LET W(6)=W(5)*R3
2340 LET O(6)=O(5)*R3
2350 LET A(6)=A(5)*R3
2360 LET D(6)=D(5)*R3
2370 LET S(6)=S(5)*R3
2380 LET I(6)=I(5)*R3
2390 LET C(6)=C(5)*R3
2400'
2410 'RECYCLE LOOP ITERATION
2420 LET CS=C(1)+C(6)
2440 IF ABS(CS-C(2))<0.001 THEN 2470
2450 LET C(2)=CS
2460 GO TO 1810
2470 '
2480 '
2490 'STREAM 7 :
2500 LET W(7)=W(5)*(1-R3)
2510 LET O(7)=O(5)*(1-R3)
2520 LET A(7)=A(5)*(1-R3)
2530 LET D(7)=D(5)*(1-R3)
2540 LET S(7)=S(5)*(1-R3)
2550 LET I(7)=I(5)*(1-R3)
2560 LET C(7)=C(5)*(1-R3)
2570 '
2580 'STREAM 8 :
2590 LET X(8)=((C(1)+I(1)+O(1))*.35+W(1))*(T8-60)
2600 LET Y(8)=((C(6)+I(6)+O(6)+D(6)+S(6))*.35+A(6)*.4+W(6))*(T8-T9)
2605 LET A(8)=A(2)-A(5)
```

BALANCE (CONTINUED)

```
2610 LET Z(8)=A(8)*.35*(T8-60)
2620 LET V(8)=(X(8)+Y(8)+Z(8))/(1300-FNL(T8))
2640 LET W(8)=W(2)-W(1)-W(6)-A(8)*7/100-V(8)
2650 IF W(8)<0 THEN 2670
2660 LET V(8)=V(8)+W(8)*(T8-60)/(1300-FNL(T8))
2670 '
2700 '
2710 'STREAM 10 :
2720 LET W(10)=W(4)-W(5)
2730 LET O(10)=O(3)-O(5)
2740 LET A(10)=A(3)-A(5)
2750 LET S(10)=S(4)-S(5)
2760 LET D(10)=D(4)-D(5)
2770 '
2830 'STREAM 11 :
2840 LET N(11)=A(10)*100*1.2/98        '20% EXCESS LIME
2850 LET W(11)=N(11)
2860 '
2870 'STREAM 12 :
2880 LET W(12)=W(10)+A(10)*18/98
2890 LET W(12)=W(12)+W(11)
2900 LET S(12)=S(10)
2910 LET D(12)=D(10)
2920 LET O(12)=O(10)
2930 LET N(12)=(1/6)*N(11)
2940 LET Y(12)=A(10)*136/98                'SULPHATE FORMED
2950 LET L(12)=W(12)+S(12)+D(12)+O(12)
2960 '
2970 'STREAM 13
2980 LET Z(13)=A(10)*28/98                 'CO2 EVOLVED
2990 '
3000 'STREAM 14
3010 LET N(14)=N(12)
3020 LET Y(14)=Y(12)
3030 LET L(14)=0.6*(N(14)+Y(14))
3040 LET LS=L(14)/L(12)
3050 LET W(14)=W(12)*LS
3060 LET O(14)=O(12)*LS
3070 LET D(14)=D(12)*LS
3080 LET S(14)=S(12)*LS
3090 '
3100 'STREAM 15 :
3110 LET W(15)=W(12)-W(14)
3120 LET O(15)=O(12)-O(14)
3130 LET D(15)=D(12)-D(14)
3140 LET S(15)=S(12)-S(14)
3150 LET L(15)=L(12)-L(14)
3160 LET K9=S(15)/L(15)
3170 '
3180 'PRODUCT STREAM 16:
```

BALANCE (CONTINUED)

```
3190 LET S(16)=S(15)
3200 LET L(16)=S(16)*100/C8
3210 LET W1=L(15)-L(16)  'WATER TO BE REMOVED
3220 LET O(16)=O(15)
3230 LET D(16)=D(15)
3240 IF W1>0 THEN 3270
3250 LET A3=0
3260 GO TO 3740
3270 '
3280 'EVAPORATOR CALCULATIONS : FEEDFORWARD,LTV-TYPE
3290 LET N3=W1/.7/V(9)  'NO.OF EFFECTS
3300 IF N3>.5 THEN 3330
3310 LET N3=1
3320 GO TO 3370
3330 IF N3<6 THEN 3360
3340 LET N3=6
3350 GO TO 3370
3360 LET N3=ABS(N3+.5)
3370 LET H(0)=FNS(T9)-FNL(T9)
3380 LET P4=FNP(T9)
3390 LET P5=14.7
3400 LET F(0)=T9
3410 FOR I=1 TO N3
3420 LET B(I)=P4-(P4-P5)/N3*I    'EVAP.PRESSURES
3430 LET F(I)=FNT(B(I))          'TEMPS.
3440 LET G(I)=F(I-1)-F(I)        'TEMP.GRADIENTS
3450 LET H(I)=FNS(F(I))-FNL(F(I)) 'LATENT HEATS
3460 NEXT I
3470 MAT R=ZER(N3+1,N3+1)
3480 MAT M=ZER(N3+1,N3+1)
3490 MAT J=ZER(N3+1,1)
3500 MAT U=ZER(N3+1,1)
3510 LET R(1,1)=-H(0)
3520 LET R(1,2)=H(1)
3530 FOR I=2 TO N3
3540 FOR J=2 TO I
3550 LET R(I,J)=G(I)
3560 NEXT J
3570 LET R(I,I)=R(I,I)-H(I-1)
3580 LET R(I,I+1)=H(I)
3590 NEXT I
3600 FOR I=1 TO (N3+1)
3610 LET R(N3+1,I)=1
3620 LET J(I,1)=G(I)*L(15)
3630 NEXT I
3640 LET J(N3+1,1)=W1
3650 MAT M=INV(R)
3660 MAT U=M*J
3670 LET U2=A3=0
3680 FOR I=1 TO N3
```

BALANCE (CONTINUED)

```
3690 LET Q(I)=500-I*300/N3     'EST.HTC.
3695 LET U2=U(I,1)*(FNL(F(I-1))-FNL(60))
3700 LET X(I)=U(I,1)*H(I-1)/(G(I)*G(I))
3710 LET A3=A3+X(I)
3720 NEXT I
3730 LET A3=A3/N3     'AVG.AREA PER EVAPORATOR.
3740 IF JS=0 THEN3770
3750 GOSUB #1
3760 '
3770 'COST ESTIMATION ROUTINE :
3780 '--------------------------
3790 '
3800 FOR I=1 TO 20
3810 LET Q(I)=B(I)=F(I)=G(I)=K(I)=H(I)=0
3820 NEXT I
3830 '(1) SLURRY HEATER :
3840 LET Q(1)=FNV(1)/6   'FT↑3  --SIZE OF EQPT.
3850 LET B(1)=(Q(1)/120)↑.6*1.35*3.6*2700  'PURCHASED COST
3860 LET F(1)=B(1)*.64    'COST OF FIELD MATERIALS
3870 LET G(1)=B(1)*.61    'COST OF FIELD LABOUR
3880 LET K(1)=B(1)*.84
3890 '(2) REACTOR WITH SCREW FEEDER :
3900 LET Q(2)=FNV(2)*T5/60    'FT↑3
3910 LET B(2)=(Q(2)/50)↑.6*1.2*1.3*4E4
3920 LET F(2)=B(2)*.65
3930 LET G(2)=B(2)*.61
3940 LET K(2)=B(2)*.76
3950 LET H(2)=FNP(T3)*FNV(2)*6.07E-5*1.25   'H.P.AT 80%EFF.
3960 '(3) FLASH :
3970 LET Q(3)=FNV(3)/30   'FT↑3
3980 LET B(3)=(Q(3)/200)↑.65*1.2*3.6*2500
3990 LET F(3)=B(3)*1.04
4000 LET G(3)=B(3)*.99
4010 LET K(3)=B(3)*1.12
4020 '(4) CENTRIFUGE #1 :
4030 LET Q(4)=FNV(4)/60/.134   'GAL./MIN.
4040 LET B(4)=(Q(4)/350)↑.6*1.3*9.5E4
4050 LET F(4)=B(4)*1.05
4060 LET G(4)=B(4)*.55
4070 LET K(4)=B(4)*.45
4080 LET H(4)=Q(4)/3.5
4090 '(5) NEUTRALIZER :
4100 LET Q(5)=FNV(10)/30
4110 IF R2>.005 THEN 4140
4120 LET Q(5)=Q(5)*R2/.005
4130 IF Q(5)<= 0 THEN 4190
4140 LET B(5)=(Q(5)/200)↑.65*1.2*3.6*3E3
4150 LET F(5)=B(5)*1.04
4160 LET G(5)=B(5)*.99
4170 LET K(5)=B(5)*1.12
```

BALANCE (CONTINUED)

```
4130 LET H(5)=(Q(5)/200)*10
4190 '(6). CENTRIFUGE #2 :
4200 LET Q(6)=(FNV(12)+(N(12)+Y(12))/125)/60/.134  'GAL/MIN.
4210 LET B(6)=(Q(6)/350)↑.6*1.3*9.5E4
4220 LET F(6)=B(6)*1.05
4230 LET G(6)=B(6)*.55
4240 LET K(6)=B(6)*.45
4250 LET H(6)=Q(6)/3.5
4260 '(7) LIMESTONE STORAGE--10 DAYS
4270 LET Q(7)=N(11)/62.5*24*10   'FT↑3
4280 LET B(7)=(Q(7)/100)↑.4*1.2E3
4290 LET F(7)=B(7)
4300 LET G(7)=B(7)*.47
4310 LET K(7)=B(7)*.4
4320 '(8) LIME SLURRY TANK -- 3 HRS.
4330 LET Q(8)=N(11)*3/1.5/62.5   'FT↑3
4340 IF Q(3)<=0 THEN 4400
4350 LET B(8)=(Q(8)/100)↑.6*1.2E3
4360 LET F(8)=B(8)*1.04
4370 LET G(8)=B(8)*.99
4380 LET K(8)=B(8)*1.12
4390 LET H(8)=(Q(8)/100)*5
4400 '(9) ACID STORAGE--10 DAYS :
4410 LET Q(9)=A(8)*(100/93)/1.8/62.5/.134*24*10   'GALS.
4420 LET B(9)=(Q(9)/1E3)↑.3*1.48*1.2E3
4430 LET F(9)=B(9)
4440 LET G(9)=B(9)*.47
4450 LET K(9)=B(9)*.4
4460 '(10) LIMESTONE CONVEYOR--50 FT.:
4470 LET Q(10)=50
4480 LET B(10)=(50)↑.65*500
4490 LET F(10)=B(10)*1.02
4500 LET G(10)=B(10)*.67
4510 LET K(10)=B(10)*.4
4520 LET H(10)=1.5
4530 '(11) OIL STORAGE -- 10 DAYS :
4540 LET Q(11)=V(8)*(1300-FNL(60))*240/1.5E5   'GALS.
4542 LET Q(11)=Q(11)+((U(1,1)-V(9))*(FNS(T9)-FNL(60))-U2)*240/1.5E5
4546 LET Q(11)=Q(11)+W(8)*(T8-60)*240/1.5E5
4547 IF Q(11)<0 THEN 4590
4550 LET B(11)=(Q(11)/1000)↑.4*1E3
4560 LET F(11)=B(11)
4570 LET G(11)=B(11)*.47
4580 LET K(11)=B(11)*.45
4590 '(12) EVAPORATORS :
4600 LET Q(12)=A3   'FT↑2 PER EVAP.
4610 IF Q(12)<=0 THEN 4660
4620 LET B(12)=(Q(12)/1E3)↑.53*1.2*1E4*N3
4630 LET F(12)=B(12)*1.15
4640 LET G(12)=B(12)*.75
```

BALANCE (CONTINUED)

```
4650 LET K(12)=B(12)*.4
4660 '(13) STEAM GENERATION PLANT -- PACKAGE UNIT :
4670 LET Q(13)=V(8)  'LB/HR.
4680 LET B(13)=(Q(13)/1E5)↑.7*1.7*1E5
4690 LET F(13)=B(13)*1.08
4700 LET G(13)=B(13)*.27
4710 LET K(13)=B(13)*.4
4720 '(14) MOYNO TO REACTOR SYSTEM :
4730 LET Q(14)=FNV(1)*FNP(T8)/.134/60  'GPM*PSI
4740 LET B(14)=(Q(14)/1000)↑.52*1E3
4750 LET F(14)=B(14)*1.14
4760 LET G(14)=B(14)*.85
4770 LET K(14)=B(14)*.4
4780 LET H(14)=FNV(1)*FNP(T8)*6.07E-5*1.25
4790 '(15) RECYCLE MOYNO :
4800 LET Q(15)=FNV(6)*FNP(T9)/.134/60   'GPM*PSI
4810 IF Q(15)=0 THEN 4870
4820 LET B(15)=(Q(15)/1000)↑.52*1E3
4830 LET F(15)=B(15)*1.14
4840 LET G(15)=B(15)*.85
4850 LET K(15)=B(15)*.4
4860 LET H(15)=FNV(6)*FNP(T9)*6.07E-5*1.25
4870 '(16) ACID FEED PUMP
4880 LET Q(16)=A(8)/62.5/1.8*(100/93)/.134/60*FNP(T8)  'GPM*PSI
4890 LET B(16)=(Q(16)/1000)↑.7*1.2E3
4900 LET F(16)=B(16)*1.14
4910 LET G(16)=B(16)*.85
4920 LET K(16)=B(16)*.4
4930 LET H(16)=A(8)*100/93/62.5/1.8*FNP(T8)*1.25*6.07E-5
4940 '(17) WATER FEED PUMP :
4950 IF W(8)<=0 THEN 5030
4960 LET Q(17)=W(8)/62.5/.134/60*FNP(T8)   'GPM*PSI
4970 LET B(17)=(Q(17)/1000)↑.52*600
4980 LET F(17)=B(17)*1.14
4990 LET G(17)=B(17)*.85
5000 LET K(17)=B(17)*.4
5010 LET H(17)=W(8)/62.5*FNP(T8)*1.25*6.07E-5
5020 GO TO 5040
5030 LET B(17)=Q(17)=F(17)=G(17)=K(17)=H(17)=0
5040 '(18) LIME SLURRY SCREW :
5050 LET Q(18)=2*N(11)/1.5/62.5/60/.134*FNP(T8)  'GPM*PSI
5060 LET B(18)=(Q(18)/1000)↑.52*600
5070 LET F(18)=B(18)*1.14
5080 LET G(18)=B(18)*.85
5090 LET K(18)=B(18)*.4
5100 LET H(18)=2*N(11)/1.5/62.5*FNP(T8)*1.25*6.07E-5
5110 '(19) ACID&WATER HEATER :
5120 LET W7=W(8)+A(8)*.4
5130 IF W(8)>0 THEN 5150
5140 LET W7=A(8)*.4
```

BALANCE (CONTINUED)

```
5150 LET Q(19)=W7*(T8-60)/250/(160-40)*LOG(160/40) 'LASTTERM=LMTD
5160 IF Q(19)<100 THEN 5190
5170 LET B(19)=(Q(19)/1000)↑.65*1.25*2.3*8E3
5180 GO TO 5200
5190 LET B(19)=(Q(19)/10)↑.15*2.0*310
5200 LET F(19)=B(19)*.71
5210 LET G(19)=B(19)*.63
5220 LET K(19)=B(19)*.87
5230 '(20) CONDENSER& COND.POT :
5240 LET Q(20)=U(N3+1,1)  'LB/HR.VAP.
5250 IF Q(20)<=0 THEN 5320
5260 LET B(20)=(Q(20)/1000)↑.65*1100
5270 LET F(20)=B(20)*.71
5280 LET G(20)=B(20)*.63
5290 LET K(20)=B(20)*.87
5300 '
5310 LET Q1=Q2=Q3=Q4=H1=0
5320 FOR I=1 TO 20
5330 LET Q1=Q1+B(I)    'TOTAL PURCHASED EQUIP.COST
5340 LET Q2=Q2+F(I)    'TOTAL FIELD MAT.COST
5350 LET Q3=Q3+G(I)    'TOTAL FIELD LABOUR COST
5360 LET Q4=Q4+K(I)    'TOTAL CONSTRN.INDIRECTS
5370 LET H1=H1+H(I)    'TOTAL H.P. REQUIREMENT
5380 LET E(I)=B(I)+F(I)+G(I)+K(I)
5390 NEXT I
5400 LET Q5=Q1+Q3+Q2+Q4      'TOTAL INSTALLED COST
5410 LET Q6=Q1*0.08      'FREIGHT & TAXES
5420 LET Q7=(Q1+Q2)*.13S    'ENGINEERING
5430 LET Q8=Q1*.19      'LAND & YARD IMPROVEMENTS
5440 LET O1=Q1*.25     'BUILDINGS
5450 LET O2=Q1*.4       'PIPING
5460 LET O3=Q1*.5       'SERVICES
5470 LET O4=Q5*.18      'CONTINGENCY & FEES
5480 LET O5=Q5+Q6+Q7+Q8+O1+O2+O3+O4    'FIXED CAPITAL INVESTMENT
5490 LET O6=O5*.15     'WORKING CAPITAL
5500 LET O7=O5+O6       'TOTAL CAPITAL INVESTMENT
5510 LET O7=O7*1.15    'ALLOWANCE FOR ESCALATION OF 15% TO END'70
5520 '
5530 'MANUFACTURING COSTS:
5540 LET Z(1)=A(8)*24*32/2000    'ACID COST
5550 LET Z(2) = N(11)*24*13/2000    'LIMESTONE
5560 LET O8=O9=0
5570 FOR I=2 TO N3+1
5580 LET O8=O8+U(I,1)
5590 NEXT I
5600 LET Z(3)=(W(8)+N(11)+V(8)-O8)/62.5/.134*24  'GALS/DAY WATER
5610 LET U2=Z(3)
5620 LET Z(3)=Z(3)*1.25/1000*.25 'WATER COST
5630 LET Z(4)=Q(11)/10*.1   'OIL COST
5640 LET J1=H1*24*.7487   'TOTAL KWH
```

BALANCE (CONTINUED)

```
5650 LET Z(5)=U1*1.25*.01 'COST
5660 LET Z(6)=3*3*24  'LABOUR
5670 LET Z(7)=3.5*24    'SUPER.
5680 LET Z(8)=.15*(Z(6)+Z(7))  'FRINGE BEN.
5690 LET Z(9)=07*.05/300   'MAINT.& REPAIR
5700 LET Z(10)=.15*Z(9)/2   'SUPPLIES,50%OF MAINT.
5710 LET Z(11)=.02*07/300   'TAXES
5720 LET Z(12)=.01*07/300 'INSURANCE
5730 LET Z(13)=.15*(Z(6)+Z(7)+Z(9)/2)   'PAYROLL OVERHEAD
5740 LET Z(14)=Z(13)   'LAB.WORK
5750 LET Z(15)=Z(13)*.5/.15   'PLANT OVERHEAD
5760 LET Z(16)=I(5)*2.5*24/2000    'RESIDUAL DISPOSAL
5770 LET R4=(R2*(1+R2)↑R3)/((1+R2)↑R3-1)   'CAP.RECOV.FACTOR
5780 LET Z(17)=R4*07/300
5790 FOR I=1 TO 17
5800 LET OS=OS+Z(I)
5810 NEXT I
5820 LET F=OS/S(15)/24
5830 IF J9=0 THEN 5870
5840 GOSUB #2
5850 GO TO 5880
5860 '
5870 RETURN
5880 END
```

Appendix I-2 Listing of Sub-routine FLOBAL

FLOBAL

```
6090 'PRINTOUT ROUTINES :
6100 '*******************
6110 PRINT "------------------------------------------------------------"
6120 PRINT
6130 PRINT
6140 PRINT"MASS BALANCE FOR ";W2;" TONS/DAY PLANT"
6150 PRINT
6151 PRINT
6152 PRINT
6153 PRINT "R1";R1,"R2";R2
6154 PRINT "R3";R3," T1";T1
6155 PRINT "T5";T5,"C8";C8
6156 PRINT
6157 PRINT "MIN COST =";F
6159 IF Z3=9 THEN 6330
6160 PRINT "   ","LBS./HR."," TONS/HR"," TONS/DAY","%COMP"
6170 PRINT
6180 PRINT
6190 FOR I=1 TO 17
6195 LET K(I)=W(I)+S(I)+D(I)+O(I)+A(I)+I(I)+C(I)+N(I)
6200 PRINT "STREAM";I
6210 PRINT "---------"
6212 IF C(I)=0 THEN 6216
6214 PRINT "CELL.",C(I),C(I)/2000,C(I)*24/2000,C(I)/K(I)*100
6216 IF W(I)=0 THEN 6220
6218 PRINT "WATER",W(I),W(I)/2000,W(I)*24/2000,W(I)*100/K(I)
6220 IF S(I)=0 THEN 6224
6222 PRINT "SUGAR",S(I),S(I)/2000,S(I)*24/2000,S(I)*100/K(I)
6224 IF A(I)=0 THEN 6228
6226 PRINT "ACID",A(I),A(I)/2000,A(I)*24/2000,A(I)/K(I)*100
6228 IF O(I)=0 THEN 6232
6230 PRINT "L.INERT",O(I),O(I)/2000,O(I)*24/2000,O(I)/K(I)*100
6232 IF I(I)=0 THEN 6236
6234 PRINT "S.INERTS",I(I),I(I)/2000,I(I)*24/2000,I(I)/K(I)*100
6236 IF D(I)=0 THEN 6240
6238 PRINT "DECOMP",D(I),D(I)/2000,D(I)*24/2000,D(I)/K(I)*100
6240 IF N(I)=0 THEN 6244
6242 PRINT "LIMEST.",N(I),N(I)/2000,N(I)*24/2000,N(I)/K(I)*100
6244 IF V(I)=0 THEN 6248
6246 PRINT "STEAM",V(I),V(I)/2000,V(I)*24/2000,V(I)/K(I)*100
6248 PRINT "TOT.MASS",K(I),K(I)/2000,K(I)*24/2000
6250 LET K(I)=0
6280 PRINT
6290 NEXT I
6310 PRINT "CO2(13)";Z(13)," DT.HEAT";Z(8)+X(8)+Y(8)
6320 PRINT
6330 PRINT "YIELD(4)";S(4)/C(1)*M1/M2,"YIELD(10)";S(10)/C(1)*M1/M2
6340 PRINT
6350 IF K9>C8 THEN 6449
6355 IF K9>C8/100 THEN 6449
```

FLOBAL (CONTINUED)

```
6360 PRINT " EVAPORATOR DATA:"
6370 PRINT
6380 PRINT " EFFECT"," TEMP","PRESS","VAP","AREA"
6390 FOR I=1 TO N3
6400 PRINT I,F(I),B(I),U(I+1,1),X(I)
6410 NEXT I
6420 PRINT
6430 PRINT "STEAM REQUIREMENT=";U(1,1);"LBS/HR"
6435 PRINT " ENTERING SUGAR CONCN.=";K9;"     LEAVING CONC.=";C8
6440 PRINT
6449 PRINT "-------------------------------------------------------------"
6450 RETURN
9999 END
```

Appendix I-3 Listing of Sub-routine FLOCOST

FLOCOST

```
6470 'COST PRINTOUT ROUTINES:
6480 PRINT
6490 PRINT
6500 PRINT
6510 PRINT "EQUIPMENT COSTS:"
6520 PRINT "****************"
6530 PRINT
6540 PRINT "CODE","SIZE","P.C.","I.C.","%P.E.C."
6550 FOR I=1 TO 20
6560 PRINT I,Q(I),B(I),E(I),B(I)*100/Q1
6570 NEXT I
6580 PRINT
6590 PRINT "TOTAL PURCHASED COST=";Q1
6600 PRINT "FIELD MATERIAL FACTOR=";Q2/Q1
6610 PRINT "FIELD LABOUR FACTOR=";Q3/Q1
6620 PRINT "MATERIAL+LABOUR FACTOR=";(Q2+Q3)/Q1
6630 PRINT "DIRECT MATERIAL FACTOR=";1+Q2/Q1
6640 PRINT "L/M FACTOR=";Q3/(Q1+Q2)
6650 PRINT "INDIRECT FACTOR=";Q4/Q1
6660 PRINT
6670 PRINT
6680 PRINT
6690 PRINT "CAPITAL COST ANALYSIS"
6700 PRINT "**********************"
6710 PRINT
6720 PRINT "ITEM";TAB(40);"COST"
6730 PRINT
6740 PRINT "INSTALLED COST";TAB(40);Q5
6750 PRINT "FREIGHT & TAXES";TAB(40);Q6
6760 PRINT "ENGINEERING";TAB(40);Q7
6770 PRINT "LAND/YARD IMPROVEMENTS";TAB(40);Q8
6780 PRINT "BUILDINGS";TAB(40);Q1
6790 PRINT "PIPING";TAB(40);Q2
6800 PRINT "SERVICES";TAB(40);Q3
6810 PRINT "CONTINGENCY & FEES";TAB(40);Q4
6820 PRINT TAB(40);"--------------"
6830 PRINT "FIXED CAPITAL INVESTMENT";TAB(40);Q5
6840 PRINT "WORKING CAPITAL";TAB(40);Q6
6850 PRINT TAB(40);"--------------"
6860 PRINT "TOTAL CAPITAL INVESTMENT";TAB(40);Q5+Q6
6870 PRINT "T.C.I.WITH 15% ESCALATION";TAB(40);Q7
6880 '
6890 PRINT
6892 PRINT "F.C.I. FACTOR =";Q5/Q1
6894 PRINT "T.C.I. FACTOR =";(Q5+Q6)/Q1
6900 PRINT
6910 PRINT
6920 PRINT "MANUFACTURING COST ANALYSIS:(PER DAY BASIS)"
6930 PRINT "****************************"
6935 PRINT
```

FLOCOST (CONTINUED)

```
6936 PRINT
6940 PRINT "ITEM";TAB(35);"COST";TAB(58);"CONTR.%"
6950 PRINT
6960 LET A$(1)="ACID"
6970 LET A$(2)="LIMESTONE"
6980 LET A$(3)="WATER"
6990 LET A$(4)="OIL"
7000 LET A$(5)="POWER"
7010 LET A$(6)="LABOUR"
7020 LET A$(7)="SUPERV."
7030 LET A$(8)="FRINGE BEN."
7040 LET A$(9)="MAINT.& REPAIR"
7050 LET A$(10)="SUPPLIES"
7060 LET A$(11)="TAXES"
7070 LET A$(12)="INSURANCE"
7080 LET A$(13)="PAYROLL OVERHEAD"
7090 LET A$(14)="LAB WORK"
7100 LET A$(15)="PLANT OVERHEAD"
7110 LET A$(16)="RESIDUE DISPOSAL"
7120 LET A$(17)="CAPITAL RECOVERY"
7130 FOR I=1 TO 17
7140 PRINT A$(I);TAB(35);Z(I);TAB(60);Z(I)*100/C9
7150 NEXT I
7160 PRINT TAB(35)"------------";TAB(58);"------------"
7170 PRINT "TOTAL MANF.COST";TAB(35);C9;TAB(60);"100"
7180 PRINT
7190 PRINT
7200 PRINT "COST OF SUGAR=";F;"$/LB.SUGAR"
7205 LET R7=C3*W2/S(15)/24
7210 PRINT "COST OF PRETREATMENT=";R7;"$/LB.SUGAR"
7220 PRINT "EXPECTED MIN.SALES PRICE= 0.02 $/LB.SUGAR"
7225 LET R8=(.02-R7-F)*S(15)*24
7230 PRINT "EXPECTED MIN.NET PROFIT PER DAY=";R8;"$/DAY"
7240 PRINT "MAX.BREAKEVEN DUMP FEE=";-R8/W2
7255 LET R6=(.03-R7-F)*S(15)*24
7260 PRINT "MAX.NET PROFIT PER DAY=";R6;"$/DAY"
7270 PRINT "MIN.BREAKEVEN DUMP FEE=";-R6/W2
7280 RETURN
7290 END
```

Appendix II Description of program inputs

Input data necessary for the computer program BALANCE:

Line No.	Statement	Data	Comments
290	Data	wt. dry feed, wt. feed water	(data in tons per day)
330	Data	wt. frns. of cellulose, solid inerts, and soluble organics in feed	
390	Data	interest rate, years of operation	
395	Let C3=	cost of pretreatment per ton of refuse	
420	Let W2=	capacity of plant	(tons of refuse per day)

All the data must be entered using the line numbers and the format as shown above.

Appendix III Results of fermentation of hydrolyzed solid wastes to ethyl alcohol, using saccharomyces cerevisiae

Run No. 1

Temperature of fermenting medium = 35°C.

pH of fermenting medium = 5.0.

Sample used for hydrolysis = ground solid wastes obtained from the U.S. Public Health Service.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	50.0	0.075	4.50
2	50.2	0.2	3.90
4	50.75	0.5	3.35
6	51.8	1.125	1.94
8	52.2	1.75	1.10
10	53.8	2.1	0.24
11	54.0	2.15	0.050

Run No. 2

Temperature of fermenting medium = 35°C.

pH of fermenting medium = 5.0.

Sample used for hydrolysis = ground solid wastes obtained from the U.S. Public Health Service.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	49.5	0.125	4.2
2	49.75	0.25	3.7
4	50.1	0.40	2.6
6	50.75	0.95	1.75
8	52	1.6	1.0
10	53.1	1.92	0.22
11	53.3	2.0	0.048

Run No. 3

Temperature of fermenting medium = 35°C.

pH of fermenting medium = 5.0.

Sample used for hydrolysis = ground solid wastes obtained from the U.S. Public Health Service.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	46.7	0.05	3.90
2	46.95	0.20	3.25
4	47.75	0.75	1.84
6	49	1.37	1.0
8	49.75	1.74	0.174
10	49.9	1.88	0.10
10.75	50.0	1.90	0.042

Run No. 4

Temperature of fermenting medium = 30°C.

pH of fermenting medium = 5.0.

Sample used for hydrolysis = ground solid wastes obtained from the U.S. Public Health Service.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	48.2	0.075	4.25
2	49.2	0.20	3.75
4	51.8	0.67	2.20
6	55.9	1.3	1.30
8	58.6	1.75	0.22
10	59.7	1.95	0.06

Run No. 5

Temperature of fermenting medium = 30°C.

pH of fermenting medium = 4.4.

Sample used for hydrolysis = ground solid wastes obtained from the U.S. Public Health Service.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	48.45	0.05	4.10
2	48.8	0.20	3.50
4	49.9	0.60	2.65
6	52.6	1.30	1.14
8	54.2	1.82	0.206
9	54.6	1.90	0.056

Run No. 6

Temperature of fermenting medium = 35°C.

pH of fermenting medium = 4.4.

Sample used for hydrolysis = ground solid wastes obtained from the U.S. Public Health Service.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	45.3	0.10	3.95
2	45.65	0.20	3.575
4	47.0	0.60	2.45
6	49.4	1.30	1.10
8	51.25	1.72	0.16
10	51.6	1.80	0.034

Run No. 7

Temperature of fermenting medium = 30°C.

pH of fermenting medium = 4.4.

Sample used for hydrolysis = 1-day household solid refuse.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	48.2	0.05	4.15
2	48.8	0.15	4.10
4	50.3	0.60	2.70
6	53.1	1.30	1.16
8	56.1	1.77	0.096
10	57.3	1.95	0.052

Run No. 8

Temperature of fermenting medium = 35°C.

pH of fermenting medium = 5.0.

Sample used for hydrolysis = Black Clawson sample containing plastics, short cellulosic fibres and organics*.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	49.1	0.05	4.10
2	49.4	0.15	3.80
4	50.3	0.40	3.30
6	51.55	0.95	2.20
8	52.6	1.50	1.10
10	53.25	1.85	0.28
12	53.6	1.95	0.043

*Sample from Black Clawson "FIBRE-CLAIM" demonstration plant at Middletown, Ohio; screen and cleaner residues - sample No. 22L5 from station 10.

Run No. 9

Temperature of fermenting medium = 35°C.

pH of fermenting medium = 5.0.

Sample used for hydrolysis = Black Clawson sample containing plastics, short cellulosic fibres and organics.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	47.4	0.05	12.6
4	47.9	0.35	11.7
8	49.1	1.3	9.80
10	50.3	2.15	7.80
12	51.7	3.05	6.10
16	54.6	4.82	2.10
20	56.0	5.8	0.084

Run No. 10

Temperature of fermenting medium = 35°C.

pH of fermenting medium = 5.0.

Sample used for hydrolysis = Black Clawson sample containing plastics, short cellulosic fibres and organics.

Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	9.2	0.1	11.80
4	9.5	0.3	11.30
8	9.9	0.7	10.50
12	10.65	1.25	9.55
16	11 7	2.1	7.2
24	14.4	4.2	2.95
28	15.35	5.2	1.00
32	15.8	5.6	0.05

Run No. 11

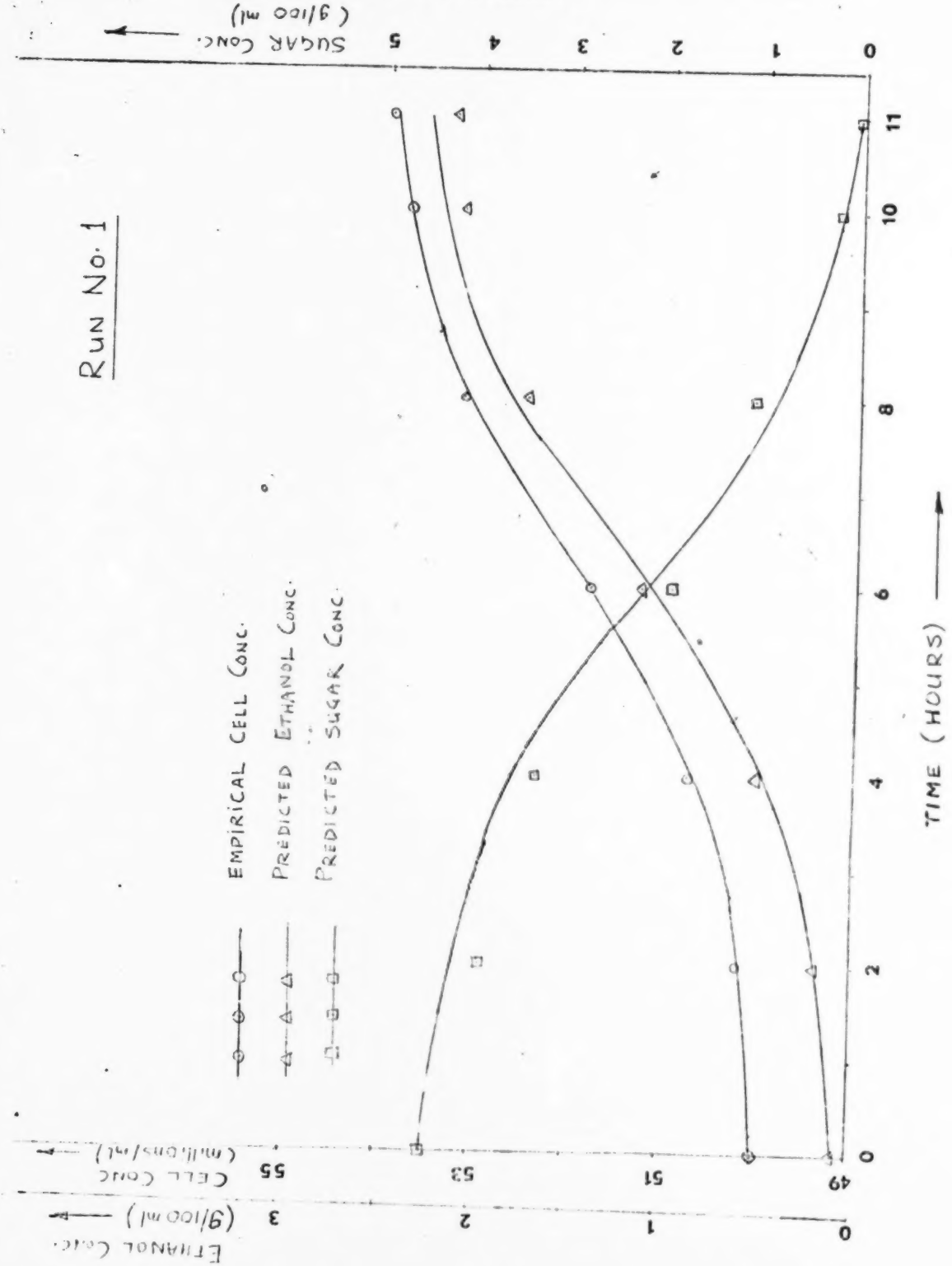
Temperature of fermenting medium = 35°C.

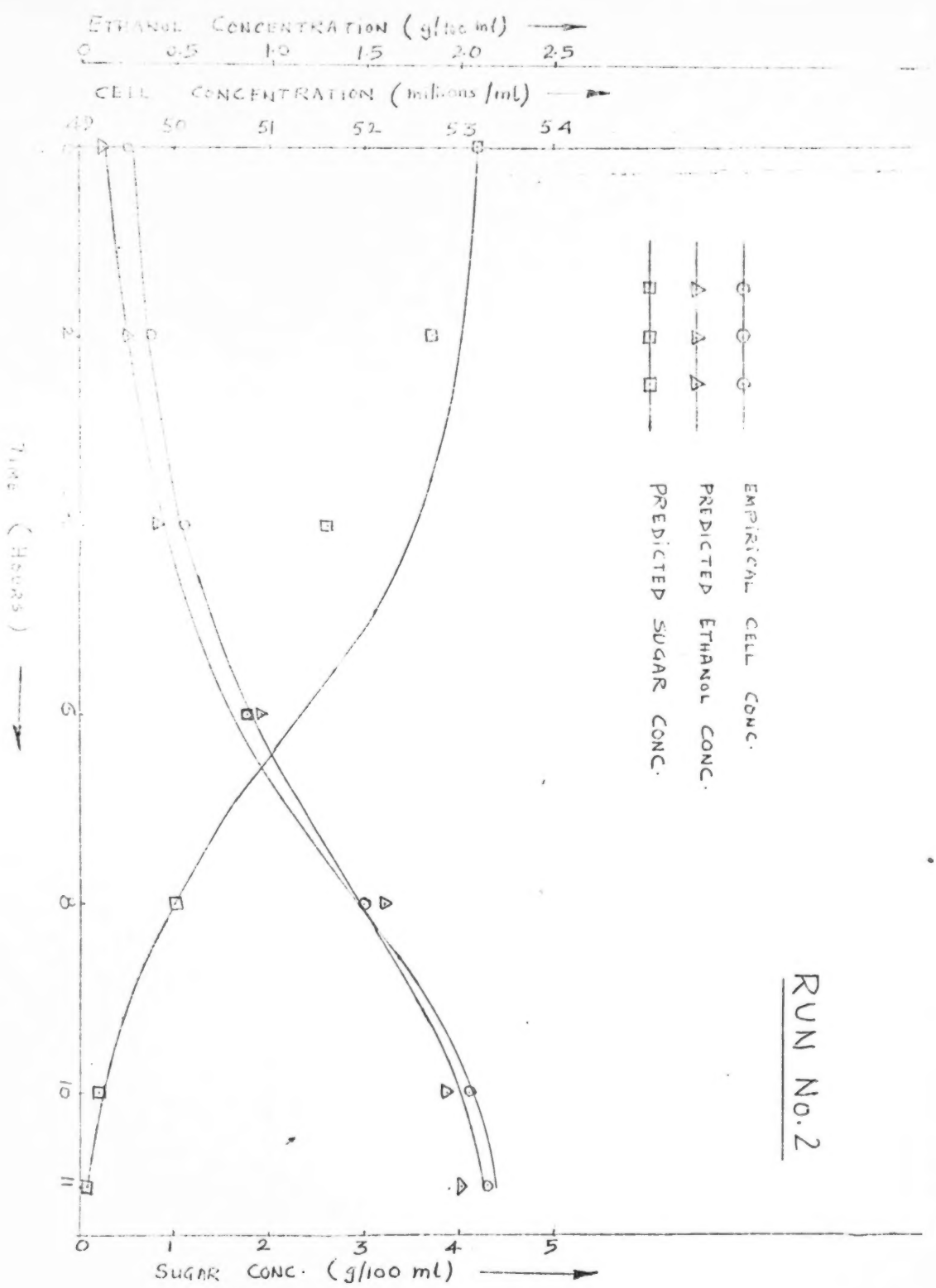
pH of fermenting medium = 5.0.

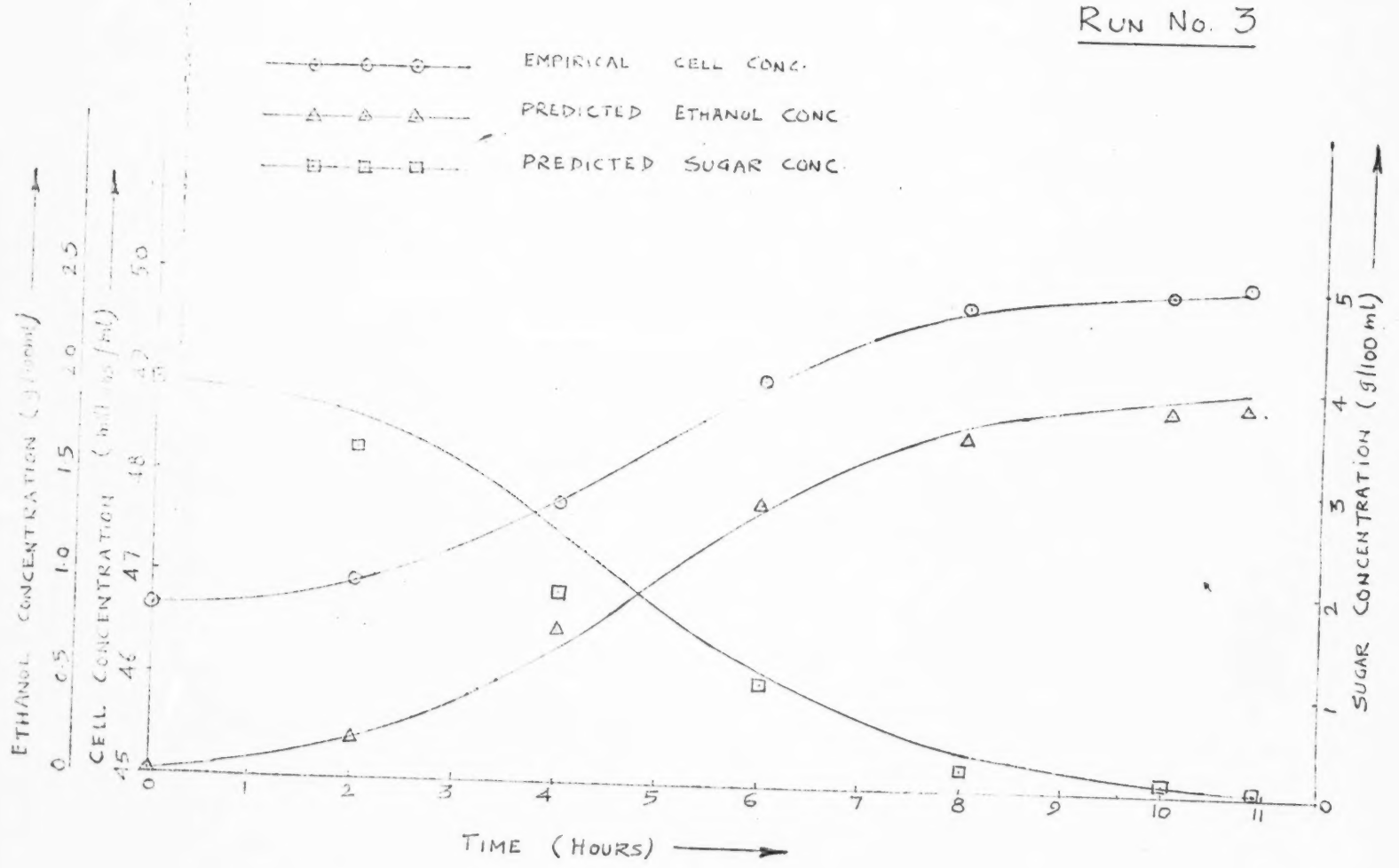
Sample used for hydrolysis - Black Clawson sample containing plastics, short cellulosic fibres and organics.

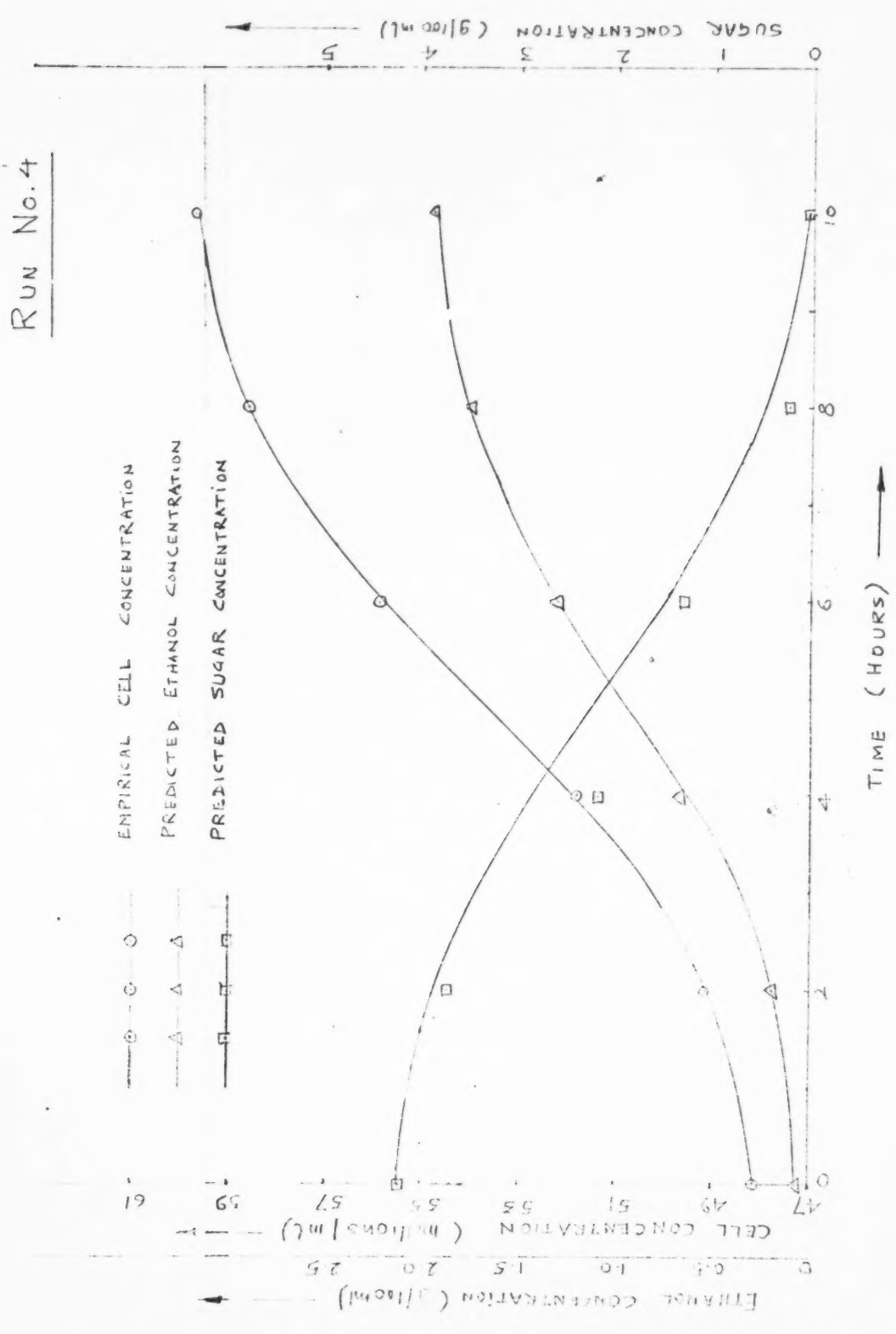
Time (Hours)	Cell Conc. (millions/ml)	Ethyl Alcohol Conc. (g/100 ml)	Sugar Conc. (g/100 ml)
0	9.8	0.075	4.050
4	10.1	0.20	3.60
8	10.6	0.37	3.10
12	11.4	0.65	2.40
20	13.3	1.50	0.75
24	13.9	1.75	0.21
26	14.1	1.85	0.032

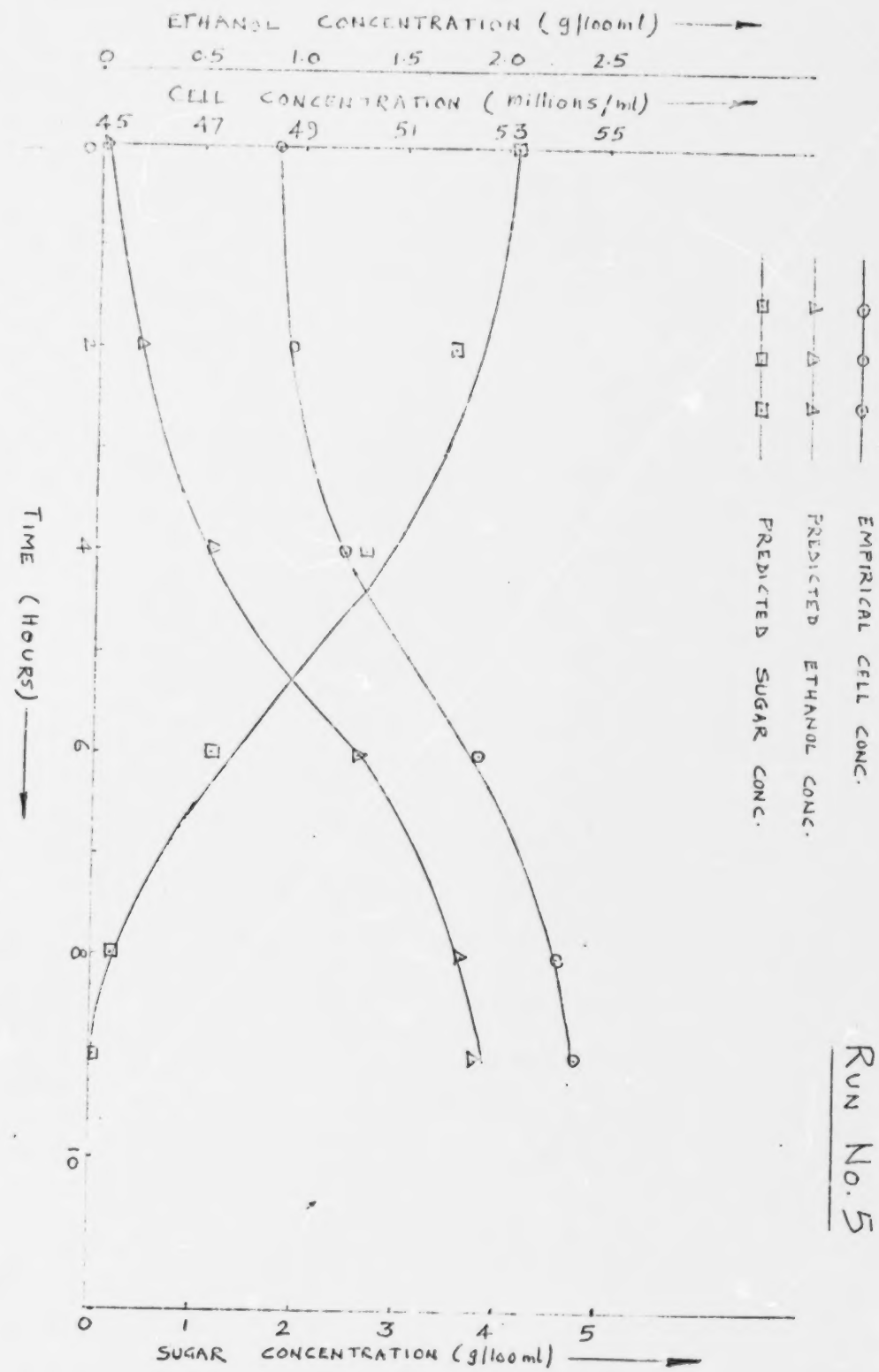
Appendix IV Summary of the experimental and the fitted curves for the microorganism concentration together with the experimental and predicted values of ethanol and sugar concentrations.



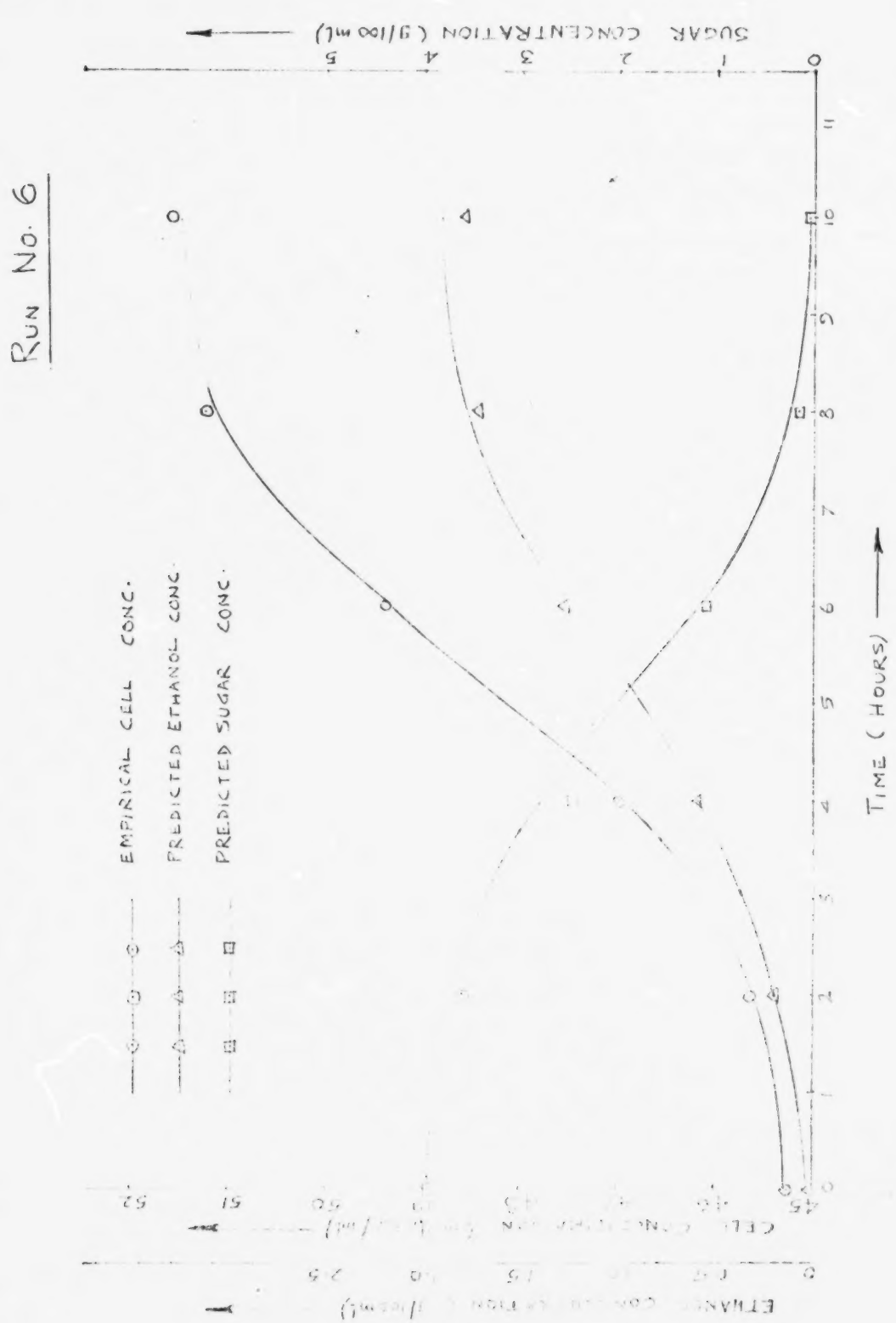

ETHANOL CONCENTRATION (g/100 ml)
0
0.5
1.0
1.5
2.0
2.5
CELL CONCENTRATION (millions/ml)
49
50
51
52
53
54
TIME (HOURS)
2
6
8
10
11
0
1
2
3
4
5
SUGAR CONC. (g/100 ml)
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.
RUN No. 2


RUN NO. 3
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC
PREDICTED SUGAR CONC
ETHANOL CONCENTRATION (g/100ml)
CELL CONCENTRATION (lakh no/ml)
SUGAR CONCENTRATION (g/100 ml)
TIME (HOURS)
0 0.5 1.0 1.5 2.0 2.5
0 15 30 45 47 18 50
0 1 2 3 4 5
0 1 2 3 4 5 6 7 8 9 10 11

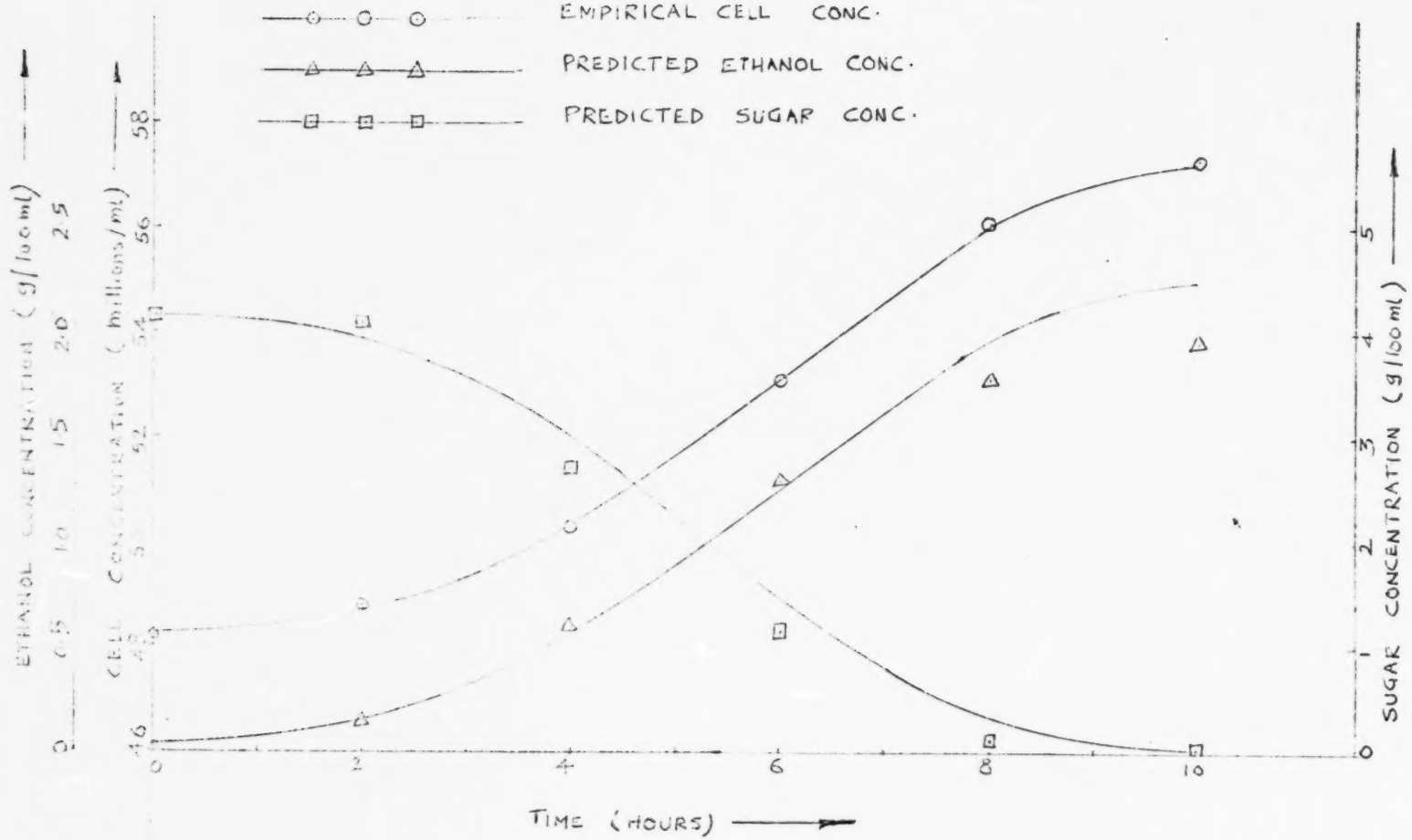

RUN No. 4
EMPIRICAL CELL CONCENTRATION
PREDICTED ETHANOL CONCENTRATION
PREDICTED SUGAR CONCENTRATION
TIME (HOURS)
CELL CONCENTRATION (millions/ml)
ETHANOL CONCENTRATION (g/100 ml)
SUGAR CONCENTRATION (g/100 ml)

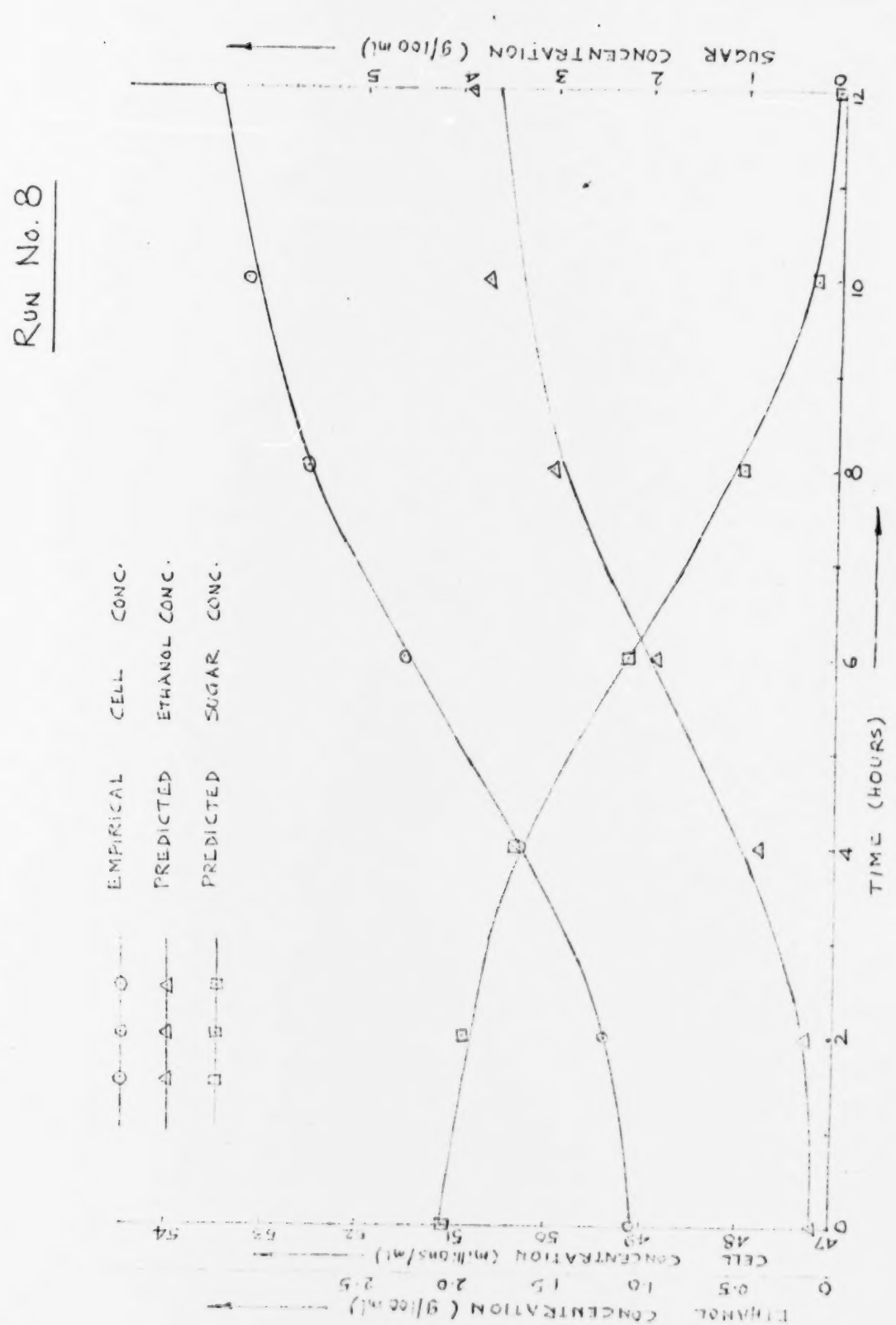

ETHANOL CONCENTRATION (g/100ml)
0
0·5
1·0
1·5
2·0
2·5
CELL CONCENTRATION (millions/ml)
45
47
49
51
53
55
TIME (HOURS)
0
2
4
6
8
10
SUGAR CONCENTRATION (g/100ml)
0
1
2
3
4
5
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.

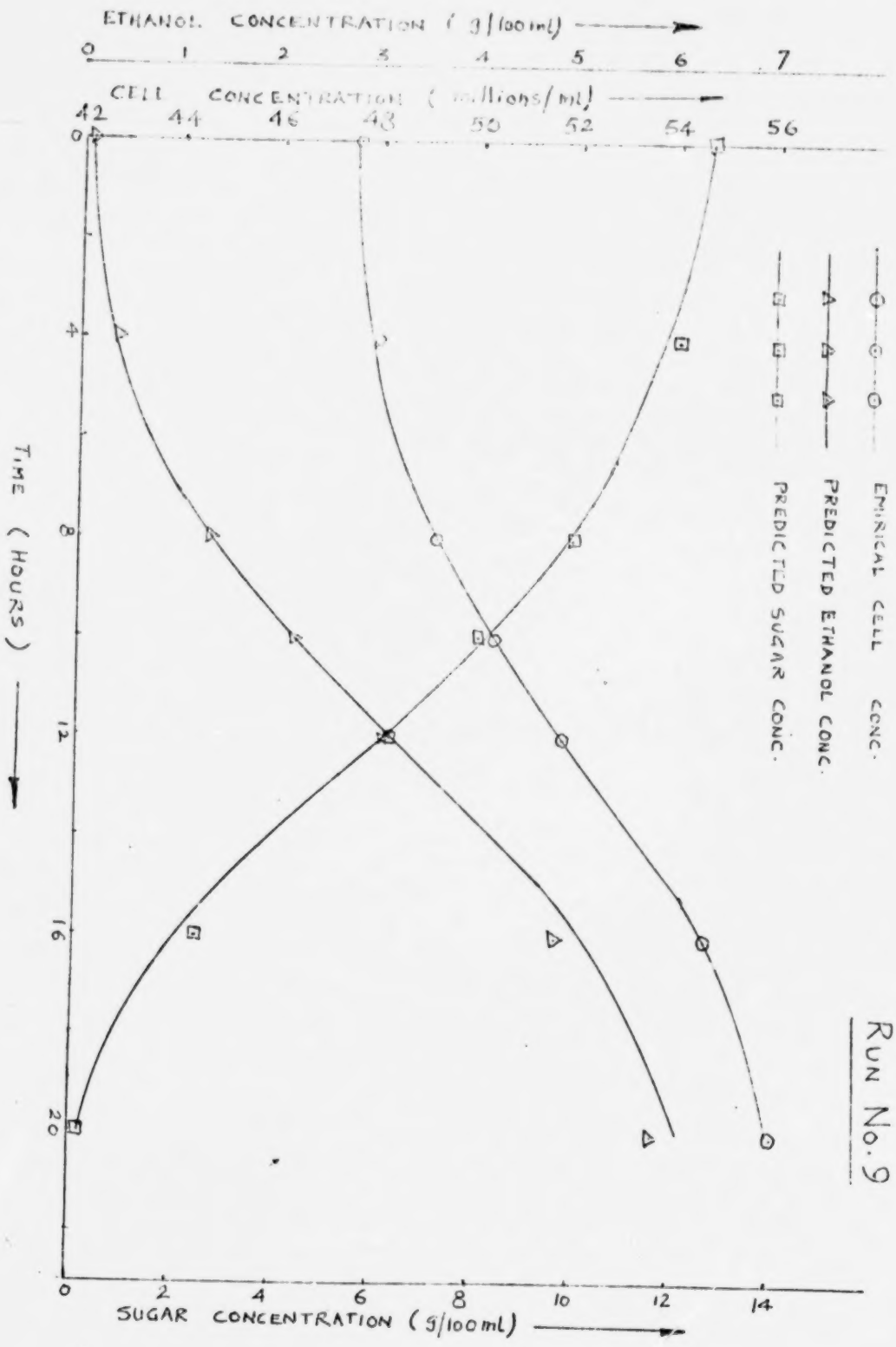
RUN NO. 5

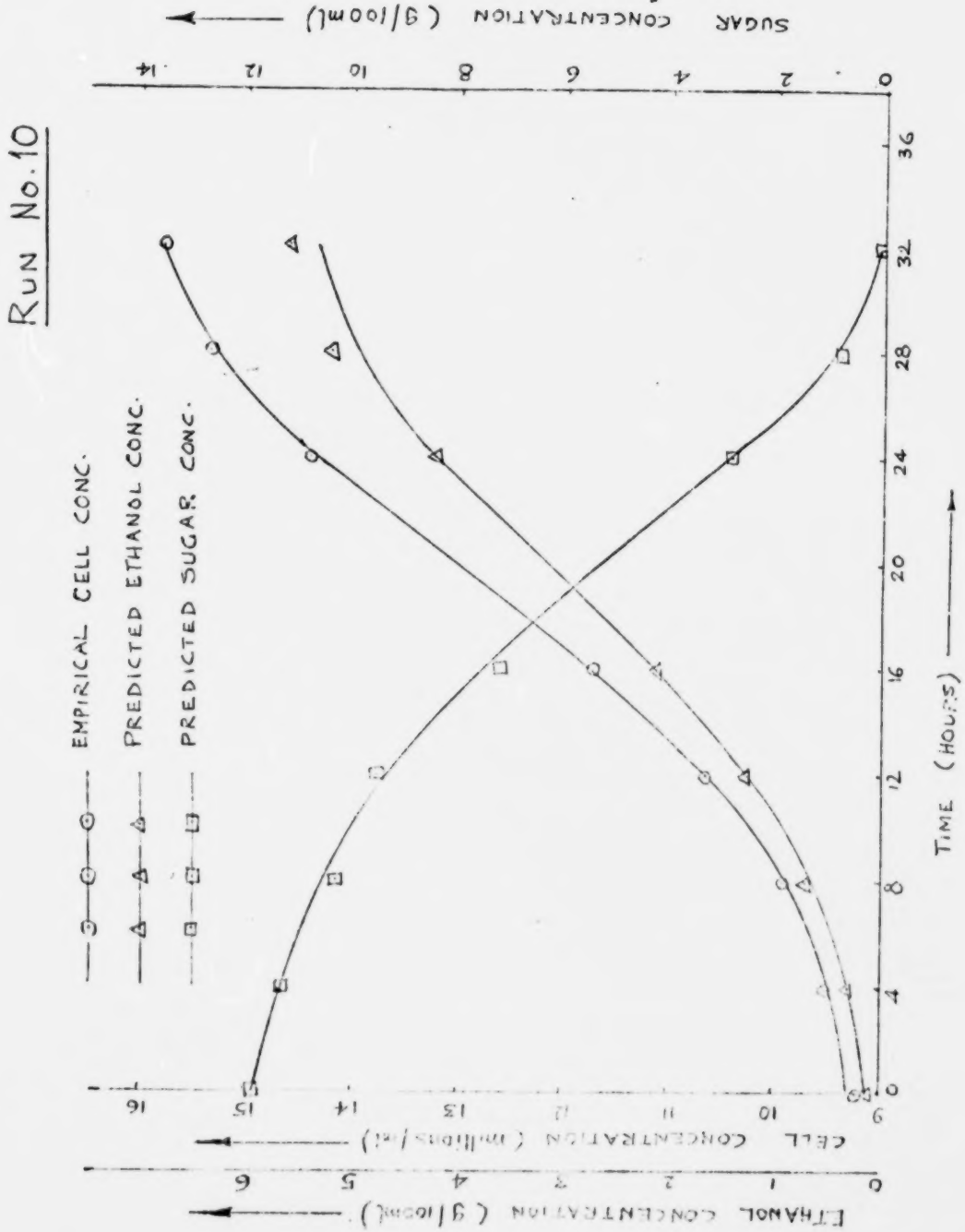

RUN NO. 6
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.
TIME (HOURS)
SUGAR CONCENTRATION (g/100 ml)

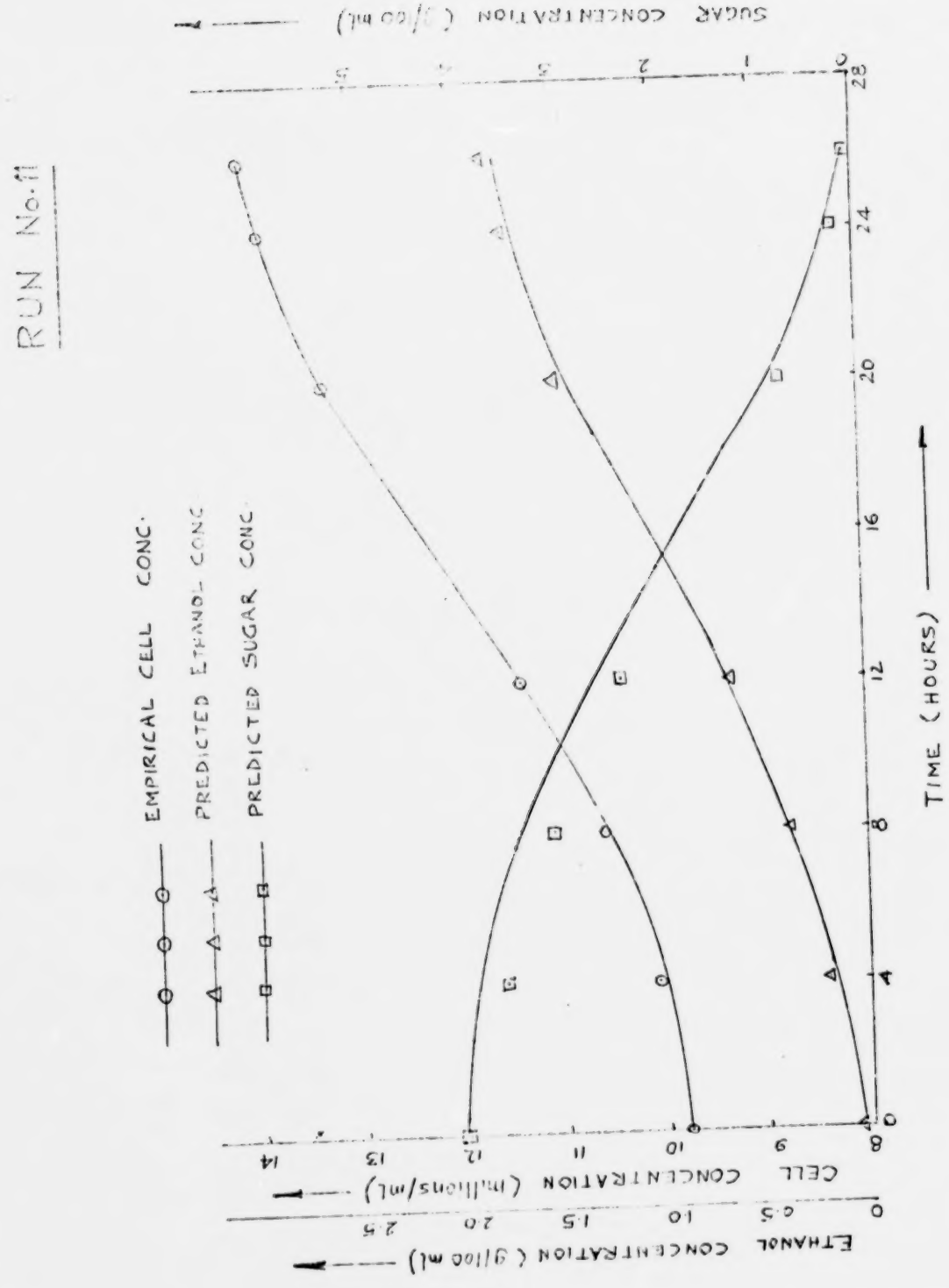
RUN NO. 7


EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.
ETHANOL CONCENTRATION (g/100ml)
CELL CONCENTRATION (millions/ml)
SUGAR CONCENTRATION (g/100ml)
TIME (HOURS)
0
2
4
6
8
10


RUN No. 8
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.
TIME (HOURS)
SUGAR CONCENTRATION (g/100 ml)
ETHANOL CONCENTRATION (g/100 ml)
CELL CONCENTRATION (mlkbns/ml)


ETHANOL CONCENTRATION (g/100ml)
0 1 2 3 4 5 6 7
CELL CONCENTRATION (millions/ml)
42 44 46 48 50 52 54 56
TIME (HOURS)
0 4 8 12 16 20
SUGAR CONCENTRATION (g/100ml)
0 2 4 6 8 10 12 14
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.
RUN No. 9


RUN No. 10
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.
ETHANOL CONCENTRATION (g/100ml)
CELL CONCENTRATION (millions/ml)
SUGAR CONCENTRATION (g/100ml)
TIME (HOURS)


RUN No. 11
EMPIRICAL CELL CONC.
PREDICTED ETHANOL CONC.
PREDICTED SUGAR CONC.
ETHANOL CONCENTRATION (g/100 ml)
CELL CONCENTRATION (millions/ml)
SUGAR CONCENTRATION ( )/100 ml)
TIME (HOURS)